PB-238 263

EVALUATION OF R AND D INVESTMENT ALTERNATIVES FOR SO_X AIR POLLUTION CONTROL PROCESSES

M. W. Kellogg Company

PREPARED FOR
National Environmental Research Center

September 1974

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TECHNICAL REPORT DATA
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1. REPORT NO. EPA-650/2-74-098	2.	PB 238 263
4. TITLE AND SUBTITLE Evaluation of R&D Investment Alternatives for SOx Air Pollution Control Processes		5. REPORT DATE September 1974 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) D. Shore, J.J. O'Donnell, and F.K. Chan		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS The M.W. Kellogg Company Research and Engineering Development Houston, Texas 77046		10. PROGRAM ELEMENT NO. 1AB013; ROAP 21ADE-029 11. CONTRACT/GRANT NO. 68-02-1308 (Task 7)
12. SPONSORING AGENCY NAME AND ADDRESS EPA, Office of Research and Development NERC-RTP, Control Systems Laboratory Research Triangle Park, NC 27711		13. TYPE OF REPORT AND PERIOD COVERED Final; 10/72-12/73 14. SPONSORING AGENCY CODE

15. SUPPLEMENTARY NOTES

16. ABSTRACT

The report presents data on sulfur oxide (SOx) emissions from five major source groups: utility plants, industrial boilers, non-ferrous smelters, sulfuric acid plants, and sulfur (Claus) plants. For all source groups studied, the bulk of the SOx emissions comes from a relatively small number of the largest plants. The report also includes evaluations of several different sulfur control systems, including stack gas scrubbing (wet limestone process and Wellman/Allied system), substitute natural gas, solvent refined coal, Lurgi gasification with a combined power cycle, and pressurized fluidized-bed combustion with a combined power cycle. Process and cost models and/or economics are presented for each system. Cost models for the stack gas scrubbing processes were applied to existing utility plants in the U.S. and the results analyzed.

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17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Air Pollution	Sulfuric Acid	Air Pollution Control	13B
Sulfur Oxides		Stationary Sources	07B
Cost Effectiveness		Industrial Boilers	14A
Boilers		Claus Plants	13A
Electric Utilities			
Smelters			11F

18. DISTRIBUTION STATEMENT Unlimited	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES 288
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE

EPA-650/2-74-098

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Contract No. 68-02-1308 (Task 7)
ROAP No. 21ADE-029
Program Element No. 1AB013

EPA Task Officer: G.J. Foley

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Prepared for

OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
WASHINGTON, D.C. 20460

September 1974

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This report has been reviewed by the Environmental Protection Agency and approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the Agency, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

EVALUATION OF R&D INVESTMENT ALTERNATIVES FOR SO_X AIR POLLUTION CONTROL PROCESSES

PART 1

TASK NO. 7 FINAL REPORT

CONTRACT NO. 68-02-1308 & CPA 70-68

by

THE M.W. KELLOGG COMPANY
RESEARCH & ENGINEERING DEVELOPMENT
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OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
WASHINGTON, D.C. 20460

SEPTEMBER, 1974

RESEARCH AND ENGINEERING DEVELOPMENT — KELLOGG

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Submitted to
ENVIRONMENTAL PROTECTION AGENCY
OFFICE OF RESEARCH & DEVELOPMENT
CONTROL SYSTEMS LABORATORY
CONTRACT NO. 68-02-1308 & CPA 70-68

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1. INTRODUCTION

The work reported herein is a technical and economic evaluation of the R&D investment alternatives for sulfur oxides pollution control methods and was performed for the Office of Research & Development, Environmental Protection Agency under Tasks 22-25, 27, Contract No. CPA 70-68 and Task 7, Contract No. 68-02-1308.

The primary objective of this work was to provide EPA with cost information for the control of sulfur oxides, which could be used to help determine regulations that can be effectively applied to the existing sulfur dioxide emissions from stationary sources. This work also attempts to provide EPA with information useful as a guide for allocating its annual development budget to produce the optimum short term and long term reduction in emissions of sulfur oxides.

The work included in this report represents Part I of a two-part study. Part I was divided into three phases:

Phase 1

To tabulate and assess information on existing sources of sulfur oxides emissions. Details of all coal-, oil- and gas-fired steam generating power plants, nonferrous smelters, coal- and oil-fired industrial boilers, acid plants and Claus plants were to be characterized and tabulated according to plant capacity, type of fuel (feed) used, age of boilers (utilities), stream factor, and geographical distribution. The results were to be stored in an accessible computer format so that reference could be made whenever necessary.

Phase 2

To study and evaluate several possible methods of sulfur dioxide emission control. A significant part of the total effort was to be allocated to the selection of processes that would be of significance in existing or potential technology.

All processes, some containing many alternate designs, were to be represented by a process and a cost model. These models relate the important process variables to the capital and operating costs of the plant. The models were to be written in such a way as to facilitate future revisions in the models as dictated by improvements in the processes. The processes were to be classified according to the following categories:

1. Stack Gas Scrubbing
   Processes representing once through ("throwaway") and regenerative types were to be evaluated. Wet limestone (throwaway) and Wellman/Allied (regenerative) were selected as the candidate processes.

2. Production of Clean and Low Sulfur Fuel
   Two different types of processes were selected for evaluation:
   - High Btu gas from coal, using a Lurgi gasification unit.
   - Highly refined coal by solvent extraction.

3. New Power Plant Designs
   Two different concepts were to be evaluated:
   - A combined-cycle power plant using low Btu gas from Lurgi gasifiers.
   - A new type of power plant design using a pressurized, fluidized-bed combustor.

After establishing these models, the cost of installing stack gas scrubbing for the existing utilities was to be investigated using the utilities emissions inventory generated from Phase 1 on a plant basis. In addition, the costs of manufacturing substitute natural gas as well as solvent refined coal were to be investigated for different parts of the country. The potential of the new power plant designs mentioned above were to be assessed and improvements, if needed, illustrated.

Phase 3

Assessment of development work in control technology and prediction of future demands of energy and chemicals. EPA requested that this be done by a modified Delphi Technique, which involved sending questionnaires to a panel of experts and reforming and expanding the questions on the basis of their answers. The results of this phase will be reported separately by EPA.

At the completion of Part I, the Environmental Protection Agency (EPA) requested that the following work be included in a second part of this study:

- Upgrade the utility boiler data base with the Federal Power Commission (FPC) Form 67 magnetic tape provided by EPA, adding to the data base:

    1. Boiler load factor
    2. Boiler fuel consumption by fuel type
    3. Boiler fuel sulfur

- Determine the costs for installing Wet-Limestone and Wellman/Allied stack gas scrubbing units on existing industrial boilers on a plant basis for all plants greater than 5 megawatts (equivalent size) and summarize the results.

- Modify the Wellman/Allied scrubbing process so that it will be applicable to acid plants, and to determine the costs for installing the regenerable scrubbing unit on existing acid plants.

- Upgrade the Claus plant data base by including the number of reaction stages per plant and investigate the feasibility of applying the regenerable scrubbing cost model to these emission sources.

- Determine the mine-mouth costs of substitute natural gas (SNG), and solvent refined coal (SRC) for different parts of the country where coal exists.

- Incorporate a cost model for the production of low and intermediate Btu gas production into the SNG cost model.

- Estimate the costs and develop cost models for shop fabrication and packaging of scrubber units for the throw-away and regenerable scrubbing processes for non-utility boilers.

The above mentioned work will be reported as Part II of the study and will be issued at a later date.

2. CONCLUSIONS AND RECOMMENDATIONS

2.1 Control of Sulfur Dioxide Emissions From Existing Sources

In every one of the 5 major SO_2 source groups studied, 75% of the emissions come from a relatively small number of the largest plants. A significant national reduction in SO_2 emissions could be achieved by directing control efforts towards these larger plants. In the case of the utility industry and industrial boilers most of these plants are concentrated in the 5 or 6 coal producing states south of the Great Lakes.

The non-ferrous smelting industry is an easy target for significant SO_2 reductions. This industry is the second largest group emitter nationally. There are only about 40 plants all told and the largest 20 emit 75% of the smelter SO_2. The cost of stack gas scrubbing controls for each of these is probably less than $30 million.

The small industrial boilers do not appear to be an optimal target for significant SO_2 reduction. About 72% of the industrial boiler population are smaller boilers of 100 MMBtu/HR or below emitting 24% of industrial boiler SO_2 emissions, and the costs for retrofitting stack gas scrubbing units for such small size boilers would be very expensive.

2.2 Stack Gas Scrubbing Costs

2.2.1 The Utility Industry

The largest 200 utility plants, those greater than 400 megawatts, are responsible for 75% of the utilities SO_2 emissions. These can be controlled by stack gas scrubbing to an overall plant emission of less than 1.2 lb SO_2/MMBtu fuel fired for total capital investments ranging from $40/plant kilowatt to $75/kilowatt. The increases in electricity costs by these controls range from 1 to 3 mils/kwh.

2.2.2 The Industrial Boilers

Typical costs were determined for a single boiler, assuming a load factor of 50%. The equivalent incremental fuel costs for clean fuel which could be absorbed as an alternative to stack gas scrubbing range from \$0.90 to \$3.30/MMBtu with decreasing boiler size. These costs could be lowered in multi-boiler plants by ducting several boilers to a common scrubbing unit. Plants with higher load factors would also have lower costs.

Assuming an average cost for high sulfur fuel of \$0.40/MMBtu, SO_2 regulations imposed on boilers of less than 100 MMBtu/hr would create demands for clean fuel at costs ranging from \$1.30 to \$3.70/MMBtu. However the 3600 boilers in this size range (almost three-fourths of all coal and oil-fired industrial boilers) account for only 24% of the U.S. industrial boiler emissions and sensible regulations would not force controls on these small boilers, unless there were particularly good local reasons.

2.3 Substitute Natural Gas Production

SNG can be produced from coal at costs ranging from \$1.20 to \$1.40/MMBtu provided coal costs are around \$3/ton and the location of the plants is not in a high construction cost area.

The HHV of the product SNG would be about 58% of the HHV of the input coal to the plant. Although this compares favorably to the efficiency of a power plant, it could be misleading. Other factors to be considered are the flexibility of the product, the relative transportation losses, the relative cost per unit of energy and the efficiency of the final product consumption for the various alternatives. The relative final efficiency of utilization is therefore not calculable without including these additional factors.

Generation of SNG for supplying clean industrial boiler fuel from low sulfur coal is probably not the best utilization of the low sulfur coal. It could probably be transported and burned directly in existing coal-fired industrial boilers at a lower overall cost.

A study of availble coals, sulfur contents, mine-mouth costs and transportation methods and costs to various areas in U.S. appears useful. Such a study should provide insights into the optimal control methods: i.e. stack gas scrubbing versus the production of clean fuels.

One point which ought to be stressed is the need to ensure that the costs of mining the low sulfur surface coal include the costs of returning the landscape to a respectable level and also include adequate compensation to the inconvenienced residents of the area.

2.4 Solvent Refined Coal Production

Solvent refined coal could undoubtably be produced for much less than SNG and the SRC plant recovers about 79% of the heat content of the coal in the products. It therefore has the advantages of cost and efficiency over SNG. It is, however, a solid and less flexible fuel, and normally contains about 1% sulfur when produced from a 4% sulfur coal (DAFB). If raw coal costs were around $3 or $4/ton, it could be produced for $0.7 to $0.9/MMBtu. It appears that as low as 0.4% sulfur and liquid fuel can be produced by slight process modification and increase in costs.

The real area for investigation appears to be the market. SRC is basically an expensive, low sulfur, ash free solid fuel not suitable for direct use with gas turbines. Since the process can also be geared towards specialized refinery type products, the question arises as to whether this would be a more worthwhile direction than the production of a solid fuel.

2.5 The Lurgi Gasifier with Combined Cycle

It does appear that the Lurgi gasifier with a combined power cycle compares favorably in both efficiency and operating costs to the conventional steam cycle power station fitted with Wellman/Allied stack gas scrubbing. It should be emphasized that the Lurgi combined cycle plant is a base loaded power station and relatively more complicated to operate than the conventional plant. There is a need for more detailed technical and cost studies of the present design. There is a very great incentive for development of better gas turbine designs with higher inlet temperatures than presently allowable.

2.6 The Fluidized Pressurized Combustor with A Combined Power Cycle

This is basically a conventional steam cycle with a small gas turbine added. Its efficiency is about the same as a conventional steam plant fitted with Wellman/Allied stack gas scrubbing. The costs presented here show it to be less expensive, but there are several areas which have not been proven even at pilot plant level. The final costs could be several million more than the figures presented here. In particular the dolomite regeneration is not proven and there is evidence that regeneration efficiency falls off with the number of regenerations. The plant is less flexible and more complicated to operate than the conventional power station. Its cost savings are uncertain at this stage and this is its biggest claim to superiority. It appears to be a less promising design than the Lurgi combined cycle.

3. MAJOR SOURCES OF SULFUR DIOXIDE EMISSION IN U.S.

3.1 Introduction

One of the main objectives of this study was to tabulate and assess information on major sources of sulfur dioxide emissions in the U.S. so that optimal ways of controlling the emissions could be evaluated. Information on five major sources of sulfur dioxide emissions have been gathered from steam generating utility plants, coal- and oil-fired industrial boilers, non-ferrous smelting industry, acid plants and sulfur (Claus) plants. The sulfur dioxide emissions studied in this report are on a national level with emphasis on optimal reduction.

The total U.S. sulfur dioxide emissions from these major sources are (in terms of sulfur emitted per year):

	Million Tons Sulfur/Year	As % of the Total
Utilities	8.742	64.7
Smelters (Lead, Zinc, Copper)	1.923	14.2
Industrial Boilers	1.761	13.0
Acid Plants	0.654	4.9
Sulfur Plants	0.437	3.2
	13.517	100.0

3.2 Data

3.2.1 Sources

The sources of data for these industries are summarized as followed:

Utilities

The utilities data were generated from a tape supplied by EPA and containing data from the National Emissions Data System, Office of Air Quality Planning and Standards (NEDS tape), and also from a tape prepared by MWK containing data from the FPC Form 67's and the National Coal Association's 1972 edition of "Steam-Electric Plant Factors" (1).

Smelters

The source of data is the report prepared by Arthur G. McKee and Company for the National Air Pollution Control Administration (2). Emissions are for the calendar year 1968.

Industrial Boilers

The NEDS tape is the sole source of data for industrial boilers. Emissions are for the calendar year 1971.

Acid Plants

The sole source of data is Chemico's report for the National Air Pollution Control Administration (3). The report contains data for acid plants in 1969.

Sulfur Plants

The NEDS tape as well as the report by Process Research Incorporated (4) were used for the sulfur plants emissions. Data are for 1971.

3.2.2 Data Quality and Comparison with Other Published Information

Sulfur emissions inventories for this report were either taken from published data or proprietary information from EPA as itemized previously. The data should not be considered as absolutely accurate or complete since they are not taken from a direct census of the emission sources. Additionally, not all of the data sources for the five industrial categories use the same base year. Comparison with the official EPA position on sulfur emissions can only be made when growth rates for the smelters and acid plants are known. Table 3.1 presents such a comparison assuming a yearly growth rate of 4% for these two industries for the purpose of updating the data to 1971 (5), thus providing a consistent time base for all five industrial categories.

It should be pointed out that there are other limitations for direct comparison of the two. First, the MWK data came from actual summations of the NEDS tapes or from other reference sources while the OAQPS data assumed a uniform emission factor, which is a simplification of the actual situation. Second, the categories of industries studied are grouped differently in the two reports; e.g., the OAQPS report grouped smelters and acid plants under industrial and chemical processes whereas this report groups them as separate categories. Third, there is no direct figure given in OAQPS report for industrial boilers and the closest comparable figure for industrial boilers in the OAQPS report is that given for other manufacturing type processes, oil and gas companies as well as steel and rolling mills.

Despite such limitations, sulfur emission statistics in this report compare favorably with the official OAQPS figures (Table 3.1) with the exception of acid plants. The OAQPS report lists the sulfuric acid production of 29 million tons in 1971 whereas the Chemico report (3) lists annual capacity of 38 million tons of 100% acid equivalent in 1968. The annual capacity is based upon 330 operating days per year equivalent, which is not an unrealistic figure for acid plants.

Table 3.1 also presents the sulfur oxides emissions from extrapolation of the InterTechnology Corporation (ITC) report for EPA (7). Note the extrapolated figures are higher than the corresponding MWK figures for both the utilities and industrial boilers. The MWK figures are probably a little low considering they are obtained from data files which were incomplete.

There are situations in the NEDS tape where information such as yearly net generation, percent of sulfur in coal, yearly fuel consumption, etc. are missing for a particular plant. Whenever possible, the MWK data for the utility industry have been updated by reference to the National Coal Association's 1972 edition of "Steam-Electric Plant Factors" and the Federal Power Commission (FPC) Form 67. However, for the case of industrial boilers, there is no comprehensive publication on sulfur emissions. A task is being undertaken by EPA to obtain a complete emissions profile for industrial and commercial boilers. Owing to the limited data available in the NEDS tape, the sulfur emissions inventory for the industrial boilers is approximately 86% complete. The emissions inventory can be updated when more information is obtained.

3.3 U. S. Sulfur Dioxide Emissions

3.3.1 Utility Plants

The utility industry burning coal, oil and gas is the largest source of the major U.S. sulfur emissions. This industry alone emitted 8.742 million tons of sulfur in 1971, representing 65% of the total U.S. sulfur emissions. Only 7.5% of the total fuel was burned in exclusively gas fired boilers which do no need stack gas cleaning.

Table 3.2 gives a national breakdown of the utilities' fuel consumption. Tables 3.3-3.6 present a statewise breakdown of the utility industry. Figures 3.1-3.8 show graphs, histograms and conclusions which can be drawn from the analysis of these statistics.

Some 880 utility plants are included in the statistical survey and their size distribution is given in Fig. 3.1. The average number of boilers per plant remains fairly constant, at about 4, for all size plants. About 95% of the utility boilers in the U.S. are less than 400 megawatts. These account for 71% of total utility boiler capacity (Fig. 3.2). A little more than 62% of the utility plant population has a plant load factor of 0.4 or above (Fig. 3.3). Only 2% of the plant population has a load factor of 0.8 or above whereas 12% of the plant population has a load factor of 0.2 or below. The average load factor for all plants considered is approximately 0.5 (Fig. 3.4).

Six states, representing 29% of the total U.S. utility capacity, contribute 54.5% of the utility sulfur emissions. These states are Ohio, Pennsylvania, Illinois, Indiana, Kentucky, and Michigan (Fig. 3.5). It is not surprising that these states are centered around the major coal fields of U.S. since 92% of the utilities sulfur emissions come from coal fired boilers.

The 22 states that individually contribute at least 1% of the total utility sulfur emissions are all in the eastern half of the country. These states, shown in Figure 3.5, represent 95.4% of the utility industry sulfur emissions. Thus, viewed on a national level, the utility industry in central and western states does not present a problem under existing conditions.

Figure 3.6 shows that significant reductions can be made in the national utility sulfur dioxide emissions by cleaning up relatively few of the largest plants. For example:

1. The coal and oil fired plants over 750 megawatts (approximately 100 plants) account for 50% of the sulfur dioxide emissions in U.S. utility industry.

2. The next 100 plants (400-750 megawatts) emit an additional 20% of the sulfur dioxide emissions.

3. The next 15% comes from approximately 180 plants in the range of 150-400 megawatts.

It is fairly clear then, viewed simply on a national level, there should be a cut-off point beyond which further sulfur dioxide reductions become increasingly more difficult to achieve. Figure 3.7 shows that the older the plant, the smaller it is and, not surprisingly, the smaller the individual boilers. It is to be expected that the cost per kilowatt of installing stack gas scrubbing units on a new boiler increases as the size of the boiler decreases. However, if the boiler is in existence already, it also becomes increasingly more difficult to retrofit a stack gas unit with increase in age and decrease in size.

The average boiler size in Figure 3.7 can be misleading, especially if the larger size utility plants have one or more very small boilers. These smaller boilers are usually operated infrequently. Therefore, the cost of retrofitting them would further increase the cost of electricity delivered.

Clearly then, a great deal of thought must be given to the regulations introduced for controlling SO_2 emissions from existing utility plants. Otherwise some very expensive and unnecessary modifications may be imposed on the utility industry.

3.3.2 Smelters (Copper, Zinc, Lead)

The sulfur dioxide emissions from the nonferrous smelting industry were analyzed based on a report for EPA by Arthur G. McKee & Company. The nonferrous smelting industry alone emits 1.923 million tons of sulfur per year or 14.2% of the total U.S. sulfur dioxide emissions and is the second largest group source of sulfur dioxide emissions. The sulfur dioxide emissions from the respective smelters are (in terms of sulfur emitted) :

Smelters	Sulfur Emissions Mton/year	% of Total Smelter Emissions
Copper	1,471	76.5
Zinc	321	21.8
Lead	131	6.7
	1,923	100.0

Three states contribute 65% of the total U.S. nonferrous smelting industry sulfur dioxide emissions. These states are Arizona, Texas and Montana. Arizona has eight copper smelters which emit 34% of U.S. smelter sulfur dioxide emissions (Fig. 3.11). Almost all of the sulfur dioxide emissions from the industry come from the western and southwestern parts of the country.

Figure 3.9 shows the number of plants and the plant size distribution. Figure 3.10 shows the relationship between the number of plants, the range of plant capacity and the cumulative percentage of the sulfur dioxide emission for a certain range of plant capacity, beginning with the largest. It can be seen that 50% of the emissions come from plants over 125 tons/ year capacity (approximately 10 plants). The next 10 plants (75-125 tons capacity) account for a further 25% of the emissions.

3.3.3 Industrial Boilers

The emissions from U.S. industrial boilers constitute the third major source of sulfur dioxide emissions based on the data available. The industrial boilers emitted 1.761 million tons of sulfur or approximately 13% of the total U.S. sulfur dioxide emissions in 1971. Industrial boilers are defined as boilers in manufacturing plants which create or change raw or unfinished materials into another form or product, including the generation of electricity (with the exception of the boilers in utility industry). The gas-fired industrial boilers which burn clean fuel are excluded from this report.

Figure 3.12 shows the number of boilers in the various ranges of boiler capacity. Tables 3.8 and 3.9 present data for coal and oil fired industrial boilers, broken down by state and size.

Based on the available data, six states, representing 50% of the total industrial boiler capacity, contribute 68% of the U.S. industrial boiler sulfur dioxide emissions. These states are Pennsylvania, Ohio, Indiana, Michigan, Illinois and Minnesota. The first five are the same states which contribute the most sulfur dioxide emissions in the utility industry. The twenty-one states which individually contribute at least 1% of the industrial boilers sulfur dioxide emissions are in the eastern half of the country (Figure 3.14).

Generally speaking, the percent load factor for the industrial boilers follows the same trend as for the utility boilers. The average percent load factor fluctuates between 0.4 and 0.6. The number of boilers that operate at a load factor below 0.2 or above 0.8 is insignificantly small.

Figure 3.15 shows that significant reductions can be made in the U.S. industrial boilers sulfur emissions by cleaning up relatively few boilers. For example:

1. The coal- and oil-fired boilers over 300 MMBtu/hr capacity (approximately 500 boilers) account for 50%

of the U.S. industrial boiler sulfur dioxide emissions.

2. The next 900 boilers (100-300 MMBtu/hr capacity) account for another 25%.

3. 1600 boilers (50-100 MMBtu/hr capacity) comprise the next 15%.

(Ideally, the data should be examined on a plant basis rather than on the individual boiler basis. However, information on number of boilers per plant, plant size, etc. was not available from the NEDS tape.)

About 72% of the U.S. industrial boiler population are small boilers of 100 MMBtu/hr or below. The load factor for these small size boilers is about 44% (Table 3.9). The cost for installing stack gas scrubbing units for such small size boilers would be very expensive. Clearly other alternatives such as burning clean fuel should be considered if regulations are to be imposed on industrial boilers.

It should be stressed that the emissions inventory for coal- and oil-fired industrial boilers in this report is incomplete owing to deficiencies in the NEDS tapes, as previously discussed. The statistics presented in Tables 3.8 and 3.9 represent all the data that were available. Approximately 93% of the boilers had information on sulfur, 91% had information on capacity, while 86% had data on the amount of fuel burned. Furthermore, the NEDS tapes have partial or no boiler emissions data for Iowa, Mississippi, New York, North Carolina, Texas, and West Virginia, and these have not been included in the statistical analysis.

3.3.4 Acid Plants

The sulfur dioxide emissions from acid plants constitute the fourth major source of sulfur emissions in the U.S. The industry emits a total of 0.654 million tons of sulfur per year or 4.9% of the total U.S. sulfur emissions. The emissions depend on the type of plant, the raw feed material and the type of product. The emissions can be in two forms, acid mist and sulfur dioxide, but both are expressed in tons of sulfur emitted.

Tables 3.10 and 3.11 show the U.S. acid plant statistics by plant size, plant type and by state. Three southern states contribute 41% of the total U.S. acid plant sulfur dioxide emissions. These states are Florida, Texas and Louisiana. These are the only states which contribute, individually, 10% or more of the U.S. acid plant sulfur dioxide emissions (Fig. 3.17). California, Illinois, and New Jersey each emit about 6%.

Figure 3.18 shows that significant reductions in the acid plant sulfur emissions can be made by cleaning up relatively few of the larger plants. For example, 50% of the emissions come from plants over 800 tons per day capacity (approximately 50 plants). The next 50 plants (450-600 tons per day capacity) account for a further 25% of the emissions.

3.3.5 Sulfur Plants

The sulfur dioxide emissions from the sulfur plants constitute the fifth major source of U.S. sulfur emissions. The industry emits a total of 0.437 million tons per year or 3.2% of the total U.S. sulfur emissions. The sulfur plants statistics are presented in Tables 3.12 and 3.13, on plant size and geographical basis, repectively.

Four states emit a total of 62% of the U.S. Sulfur Plant emissions. These states are Texas, California, Mississippi and Wyoming. Texas alone emits 31% (Fig. 3.20). The major zone centers in the south with six states emitting 50% of the total sulfur plant SO_2 emissions.

Figure 3.19 shows the number of plants and the plant size distribution. Figure 3.21 shows a significant reduction in the U.S. sulfur plant sulfur emissions would be achieved by cleaning up relatively few plants. Sulfur plants over the size of 300 tons per day capacity (18 plants), account for approximately 50% of the sulfur plant emissions. The next 26 plants (100-300 tons per day of capacity) comprise the next 25%. However, the next 67 plants emit only 15% of the sulfur plant emissions.

3.4 Summary

In every one of the five major sulfur dioxide source groups, the majority (about 75%) of the emissions come from a relatively small number of the largest plants. Significant national reduction in sulfur dioxide emissions could be achieved by directing control efforts towards these larger plants. The costs of controlling sulfur dioxide emissions for these major sources will be analyzed and assessed in subsequent sections of the report.

TABLE 3.1

COMPARISON OF SULFUR DIOXIDE EMISSIONS BETWEEN THREE DIFFERENT SOURCES

	SO_2 in 10^6 TONS		
	OAQPS[1]	ITC[2]	MWK[3]
Utilities	20.1	20.5	17.5
Industrial Boilers	4.2	6.8	4.1[4]
Smelters	4.0	U	4.3[5]
Acid Plant	0.6	U	1.4[5]
Sulfur (Claus) Plant	U	U	1.0
Total	28.9	27.3	27.7

NOTES:

U = Unavailable

1. "Data File of National Emissions 1971", Office of Air Quality Planning and Standards (OAQPS), U.S. Environmental Protection Agency.
2. Extrapolation from InterTechnology Corporation report on "Energy Scenario Consumption and Consideration" as reported by G.T. Rochelle in "SO_2 Control Technology For Combustion Sources", Task 6 Final Report, EPA contract 68-02-1308.
3. Based on M.W. Kellogg summation of NED tapes and other reference materials specified previously.
4. Adjusted from MWK figure of 3.5 x 10^6 tons which represents 86% of the SO_2 emissions from Industrial Boilers.
5. Prorated to 1971 assuming a yearly growth rate of 4.0%. The base year for acid plant is 1969. The base year for smelters is 1968.

TABLE 3.2

UTILITIES FUEL CONSUMPTION AND SULFUR EMISSION FOR 1971

PLANT SIZE (MW)	FUEL BURNED BY UTILITIES IN 1971		
	COAL (%)	OIL (%)	GAS (%)
0-100	2.11	0.65	2.89
101-200	3.02	1.57	3.06
201-400	7.67	3.07	4.62
401-600	9.29	3.52	4.66
601-800	6.37	2.33	3.03
801-1000	4.58	1.69	5.15
1001-1200	7.03	0.68	1.21
1201-1400	4.02	0.60	2.03
1401-1600	1.91	0.95	1.09
1601-3000	8.59	1.05	1.54
	54.60	16.11	29.29

Total sulfur emissions for year were 8742 M tons.

TABLE 3.3

STATEWISE DISTRIBUTION OF FUEL BURNED BY UTILITIES IN 1971

STATE	COAL (%)	OIL (%)	GAS (%)
ALABAMA	2.84	0.20	0.14
ALASKA	0.00	0.00	0.00
ARIZONA	0.06	0.02	0.54
ARKANSAS	0.00	0.12	0.65
CALIFORNIA	0.00	1.53	4.43
COLORADO	0.59	0.02	0.34
CONNECTICUT	0.25	0.97	0.00
DELAWARE	0.27	0.07	0.03
D. C.	0.05	0.14	0.00
FLORIDA	0.75	2.17	1.90
GEROGIA	1.54	0.07	0.45
HAWAII	0.00	0.00	0.00
IDAHO	0.00	0.00	0.00
ILLINOIS	4.31	0.31	0.72
INDIANA	3.72	0.02	0.21
IOWA	0.66	0.00	0.48
KANSAS	0.07	0.01	1.21
KENTUCKY	3.28	0.01	0.07
LOUISIANA	0.00	0.02	2.73
MAINE	0.00	0.23	0.00
MARYLAND	0.98	0.60	0.00
MASSACHUSETTS	0.05	1.96	0.06
MICHIGAN	3.66	0.34	0.39
MINNESOTA	0.84	0.03	0.39
MISSISSIPPI	0.09	0.04	0.71
MISSOURI	1.90	0.01	0.47
MONTANA	0.08	0.00	0.01
NEBRASKA	0.16	0.00	0.34
NEVADA	0.22	0.01	0.28
NEW HAMPSHIRE	0.18	0.12	0.00
NEW JERSEY	0.63	1.54	0.21
NEW MEXICO	0.90	0.01	0.38
NEW YORK	1.47	3.13	0.61
N. CAROLINA	3.21	0.01	0.08
NORTH DAKOTA	0.45	0.00	0.00
OHIO	6.31	0.02	0.09
OKLAHOMA	0.10	0.00	1.84
OREGON	0.00	0.00	0.01
PENNSYLVANIA	5.72	1.08	0.06
RHODE ISLAND	0.00	0.12	0.02
S. CAROLINA	0.82	0.06	0.26
SOUTH DAKOTA	0.04	0.01	0.02
TENNESSEE	2.40	0.00	0.14
TEXAS	0.00	0.01	8.80
UTAH	0.08	0.09	0.01
VERMONT	0.01	0.00	0.00
VIRGINIA	1.06	0.97	0.01
WASHINGTON	0.00	0.00	0.00
W. VIRGINIA	2.81	0.02	0.00
WISCONSIN	1.64	0.03	0.18
WYOMING	0.41	0.00	0.02
	54.60	16.11	29.29

TABLE 3.4
U.S. UTILITY INDUSTRY (1971) - CAPACITY DISTRIBUTION
(All Plants)

STATE	NO. OF PLANTS	AVGE SIZE (MKW)	AVGE PCT LOAD FACTOR	AVGE BOILER SIZE (MKW)	AVGE NO. BLRS/PLANT	TOTAL CAPACITY (MKW)	PCT OF TOT U.S.
ALABAMA	14.	649.3	55.31	178.6	3.2	9090.	3.28
ALASKA	0.	0.0	0.0	0.0	0.0	0.	0.0
ARIZONA	11.	183.5	38.85	82.6	2.4	2019.	0.73
ARKANSAS	8.	283.8	39.16	158.6	2.6	2270.	0.82
CALIFORNIA	33.	577.9	44.05	122.5	5.3	19070.	6.89
COLORADO	19.	109.6	45.62	69.3	4.4	2082.	0.75
CONNECTICUT	12.	258.1	38.18	38.5	8.6	3097.	1.12
DELAWARE	4.	224.3	53.57	78.1	3.7	897.	0.32
D. C.	2.	412.0	25.35	28.4	14.5	824.	0.30
FLORIDA	36.	314.2	49.08	130.3	3.2	11310.	4.09
GEORGIA	12.	481.1	53.94	155.8	3.4	5773.	2.09
HAWAII	0.	0.0	0.0	0.0	0.0	0.	0.0
IDAHO	0.	0.0	0.0	0.0	0.0	0.	0.0
3 ILLINOIS	35.	370.4	46.64	89.3	5.8	14444.	5.22
4 INDIANA	30.	354.6	46.02	98.9	4.7	10637.	3.84
IOWA	33.	81.6	38.36	35.0	4.0	2694.	0.97
KANSAS	32.	115.3	37.33	43.7	6.3	3691.	1.33
5 KENTUCKY	16.	571.1	47.84	129.1	4.0	9138.	3.30
LOUISIANA	19.	422.7	40.01	134.1	4.3	8032.	2.90
MAINE	5.	92.2	49.89	32.2	4.3	461.	0.17
MARYLAND	11.	427.5	47.45	141.3	3.3	4702.	1.70
MASSACHUSETT	26.	199.3	45.36	74.3	4.0	5181.	1.87
6 MICHIGAN	35.	298.4	48.21	81.7	5.6	10445.	3.77
MINNESOTA	37.	90.0	33.62	44.1	4.3	3330.	1.20
MISSISSIPPI	12.	240.0	44.05	153.7	2.6	2880.	1.04
MISSOURI	27.	254.4	39.95	113.7	3.6	6869.	2.48
MONTANA	4.	74.8	32.39	97.3	1.0	299.	0.11
NEBRASKA	17.	89.5	32.12	0.0	0.0	1521.	0.55
NEVADA	6.	415.2	51.51	207.6	2.0	2491.	0.90
NEW HAMPSHIR	5.	139.6	46.81	106.3	3.0	698.	0.25
NEW JERSEY	16.	418.4	52.59	99.6	4.5	6695.	2.42
NEW MEXICO	15.	233.3	52.70	110.3	3.1	3499.	1.26
NEW YORK	33.	444.3	45.18	81.4	4.3	14662.	5.30
N. CAROLINA	14.	532.6	60.41	144.4	4.3	7456.	2.69
NORTH DAKOTA	14.	63.6	39.69	50.0	2.0	891.	0.32
1 OHIO	47.	374.3	41.07	111.0	5.1	17592.	6.35
OKLAHOMA	16.	284.6	46.70	30.7	4.7	4554.	1.65
OREGON	1.	36.0	1.59	5.1	7.0	36.	0.01
2 PENNSYLVANIA	40.	454.1	53.19	98.5	5.3	18166.	6.56
RHODE ISLAND	4.	80.8	41.79	26.0	4.0	323.	0.12
S. CAROLINA	13.	247.7	51.68	78.6	3.3	3220.	1.16
SOUTH DAKOTA	7.	30.7	33.29	16.4	2.8	215.	0.08
TENNESSEE	7.	1063.3	49.18	0.0	0.0	7443.	2.69
TEXAS	80.	346.5	43.48	119.9	3.5	27720.	10.01
UTAH	8.	63.4	37.65	80.8	1.0	507.	0.18
VERMONT	2.	17.0	19.88	0.0	0.0	34.	0.01
VIRGINIA	12.	432.1	51.93	141.4	3.3	5185.	1.87
WASHINGTON	0.	0.0	0.0	0.0	0.0	0.	0.0
W. VIRGINIA	13.	634.0	57.76	311.0	2.1	8242.	2.98
WISCONSIN	25.	209.0	46.40	60.2	4.2	5224.	1.89
WYOMING	8.	152.3	50.10	91.5	2.2	1218.	0.44
	880.					276827.	100.00

The six states of Ohio, Pennsylvania Illinois, Indiana, Kentucky and Michigan, which contribute 54.5% of the total SO_2 emissions from the U.S. utility industry, represent 29.0% of the U.S. total generating capacity.

TABLE 3.5
U.S. UTILITY INDUSTRY STATEWISE SO_2 EMISSIONS (1971)
(All Plants Except Gas-Fired Only)

STATE	NO. OF PLANTS	AVGE SIZE (MKW)	AVGE PCT LOAD FACTOR	AVGE BOILER SIZE (MKW)	AVGE NO. BLRS/PLANT	SULFUR EMITTED PER YEAR (PCT OF TOTAL U.S.) COAL	OIL	BOTH
ALABAMA	14.	649.3	55.31	178.6	3.2	4.55	0.08	4.63
ALASKA	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ARIZONA	8.	216.3	41.57	92.4	2.6	0.02	0.01	0.02
ARKANSAS	8.	283.8	39.16	158.6	2.6	0.0	0.09	0.09
CALIFORNIA	32.	592.8	44.73	123.7	5.4	0.0	0.26	0.26
COLORADO	17.	121.5	46.62	69.3	4.4	0.23	0.01	0.23
CONNECTICUT	12.	258.1	38.18	38.5	8.6	0.39	0.65	1.04
DELAWARE	4.	224.3	53.57	78.1	3.7	0.39	0.02	0.41
D. C.	2.	412.0	25.35	28.4	14.5	0.04	0.06	0.09
FLORIDA	35.	321.0	49.70	130.3	3.2	1.62	1.38	3.01
GEORGIA	12.	481.1	53.94	155.8	3.4	1.64	0.07	1.71
HAWAII	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0
IDAHO	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0
3 ILLINOIS	37.	388.8	48.73	95.0	5.6	9.11	0.12	9.23
4 INDIANA	30.	354.6	46.02	98.9	4.7	8.34	0.00	8.34
IOWA	33.	81.6	38.36	35.0	4.0	1.40	0.00	1.40
KANSAS	18.	163.9	43.18	44.2	6.6	0.15	0.01	0.16
5 KENTUCKY	16.	571.1	47.84	129.1	4.0	7.16	0.00	7.16
LOUISIANA	5.	363.4	46.00	95.6	3.3	0.0	0.01	0.01
MAINE	5.	92.2	49.89	32.2	4.3	0.0	0.20	0.20
MARYLAND	11.	427.5	47.45	141.3	3.3	1.22	0.32	1.54
MASSACHUSETT	26.	199.3	45.36	74.3	4.0	0.04	1.11	1.16
6 MICHIGAN	35.	298.4	48.21	81.7	5.6	6.09	0.22	6.31
MINNESOTA	35.	94.2	35.46	44.1	4.8	1.28	0.02	1.30
MISSISSIPPI	9.	310.2	49.04	153.7	2.6	0.14	0.03	0.16
MISSOURI	24.	284.3	41.01	113.7	3.6	4.79	0.00	4.79
MONTANA	2.	111.5	59.10	111.5	1.0	0.05	0.0	0.05
NEBRASKA	10.	129.4	34.20	0.0	0.0	0.26	0.00	0.26
NEVADA	6.	415.2	51.51	207.6	2.0	0.07	0.00	0.07
NEW HAMPSHIR	5.	139.6	46.81	106.3	3.0	0.24	0.10	0.34
NEW JERSEY	16.	418.4	52.59	99.6	4.5	0.73	0.38	1.11
NEW MEXICO	7.	407.4	64.93	153.2	3.4	0.53	0.01	0.54
NEW YORK	31.	469.2	47.59	81.4	4.3	1.81	1.51	3.32
N. CAROLINA	14.	532.6	60.41	144.4	4.3	2.30	0.00	2.30
NORTH DAKOTA	14.	63.6	39.69	50.0	2.0	0.41	0.00	0.41
1 OHIO	46.	377.7	41.38	114.0	5.1	13.52	0.00	13.53
OKLAHOMA	5.	451.8	50.77	72.9	6.2	0.07	0.00	0.08
OREGON	1.	36.0	1.59	5.1	7.0	0.0	0.00	0.00
2 PENNSYLVANIA	40.	454.1	53.19	98.5	5.3	9.60	0.37	9.96
RHODE ISLAND	4.	80.8	41.79	26.0	4.0	0.0	0.10	0.10
S. CAROLINA	13.	247.7	51.68	78.6	3.3	0.64	0.04	0.68
SOUTH DAKOTA	7.	30.7	33.29	16.4	2.8	0.03	0.01	0.04
TENNESSEE	7.	1063.3	49.18	0.0	0.0	4.49	0.0	4.49
TEXAS	20.	443.1	42.77	127.4	4.3	0.00	0.00	0.00
UTAH	8.	63.4	37.65	80.8	1.0	0.04	0.03	0.06
VERMONT	2.	17.0	19.88	0.0	0.0	0.02	0.00	0.02
VIRGINIA	12.	432.1	51.93	141.4	3.3	0.95	0.93	1.88
WASHINGTON	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0
W. VIRGINIA	13.	634.0	57.76	311.0	2.1	4.27	0.01	4.28
WISCONSIN	25.	209.0	46.40	60.2	4.2	3.00	0.00	3.01
WYOMING	8.	152.3	50.10	91.5	2.2	0.26	0.00	0.26
	744.					91.89	8.17	100.00

Six states contribute 54.5% of the total SO_2 emissions from the U.S. utility industry:

1 Ohio	13.5%
2 Pennsylvania	10.0%
3 Illinois	9.2%
4 Indiana	8.3%
5 Kentucky	7.2%
6 Michigan	6.3%
	54.5%

92% of the U.S. utilities' SO_2 emissions come from burning coal.

TABLE 3.6

U.S. UTILITY STATISTICS BY PLANT SIZE

A) ALL PLANTS

SIZE (MKW)	NO.	AVGE SIZE (MKW)	AVGE AGE (YRS)	AVGE PCT LOAD FACTOR	BOILER AVGES SIZE (MKW)	NO. PER PLANT
0- 100	351.	39.5	25.9	35.13	17.8	3.4
101- 200	142.	145.0	21.2	46.39	35.6	4.1
201- 400	150.	290.0	19.3	49.77	60.7	4.8
401- 600	92.	494.3	14.4	55.62	103.7	4.8
601- 800	47.	683.6	11.4	55.22	186.1	3.7
801-1000	35.	898.1	10.2	53.11	218.9	4.1
1001-1200	21.	1106.4	11.9	60.83	234.4	4.5
1201-1400	16.	1272.3	9.5	50.87	261.5	4.9
1401-1600	7.	1534.0	9.3	56.53	275.1	5.6
1601-3000	19.	1867.8	6.1	50.37	413.4	4.5
	880.					

B) ALL PLANTS EXCEPT GAS FIRE ONLY

SIZE (MKW)	NO.	AVGE SIZE (MKW)	AVGE AGE (YRS)	AVGE PCT LOAD FACTOR	SULFUR EMITTED PER YEAR (PCT OF TOTAL U.S.) FROM COAL	FROM OIL	FROM BOTH
0- 100	277.	39.4	26.1	35.74	3.50	0.35	3.85
101- 200	124.	145.7	21.8	47.99	4.20	0.87	5.07
201- 400	133.	292.1	19.8	51.42	13.81	1.63	15.44
401- 600	79.	493.5	15.1	56.79	15.75	1.82	17.57
601- 800	43.	683.5	11.8	55.28	10.05	1.18	11.24
801-1000	29.	893.2	10.6	52.94	6.73	0.71	7.43
1001-1200	20.	1105.0	12.2	61.49	10.49	0.40	10.89
1201-1400	14.	1275.0	9.7	49.12	8.11	0.37	8.48
1401-1600	6.	1531.3	10.1	54.28	4.88	0.56	5.44
1601-3000	19.	1867.8	6.1	50.37	14.37	0.27	14.63
	744.				91.89	8.17	100.00

TABLE 3.7

U.S. SMELTERS SO_2 EMISSIONS

(Lead, Zinc, Copper)

RANGE TONS/YR	NO. OF PLANTS	TOTAL CAPACITY TONS/YR	AVG. CAP. TONS/YR	% OF SMELTER CAPACITY	SULFUR MTON/YR	% OF TOTAL SULFUR
0-50	7	250	35.7	6.91	141.2	7.34
51-100	15	1168	77.9	32.32	542.2	28.20
101-250	9	1041	115.7	28.88	835.1	43.42
151-200	2	355	177.5	9.82	182.2	9.47
201-250	1	215	215	5.95	7.2	0.37
251-300	1	252	252	6.97	80.8	4.20
301-350	1	332	332	9.15	134.3	6.98
	36				1923.0	100.0%

TABLE 3.8

INDUSTRIAL BOILERS, COAL AND OIL FIRED
BOILER STATISTICS BY CAPACITY

RANGE	# OF BOILERS	AVERAGE CAPACITY	TOTAL CAPACITY	% OF U.S. CAPACITY	% OF TOTAL FUEL BURNED	% OF TOTAL SULFUR	PERCENT LOAD FACTOR
0- 50	2612.	22.8	50023.	9.212	17.221	14.372	44.977
51- 100	962.	74.9	72054.	13.270	11.640	10.511	42.497
101- 150	488.	126.4	61661.	11.356	11.410	10.501	40.866
151- 200	296.	175.4	51912.	9.560	9.194	7.173	40.246
201- 250	145.	227.4	32977.	6.073	6.152	4.385	41.800
251- 300	124.	276.1	34232.	6.304	5.400	5.296	47.622
301- 350	53.	328.4	17406.	3.206	2.586	2.403	41.956
351- 400	62.	381.8	23670.	4.359	2.742	2.772	36.384
401- 450	18.	438.2	7887.	1.452	0.517	0.395	19.238
451- 500	38.	476.6	18111.	3.335	4.271	4.699	54.561
501- 550	20.	520.3	10406.	1.916	1.615	1.803	46.219
551- 600	26.	583.4	15168.	2.793	2.657	3.431	53.063
601- 650	15.	631.8	9477.	1.745	1.369	1.197	54.798
651- 700	11.	680.6	7487.	1.379	2.926	3.034	56.654
701- 750	7.	742.4	5197.	0.957	0.931	1.211	20.151
751- 800	14.	779.6	10914.	2.010	1.635	1.572	49.246
801- 850	6.	824.7	4948.	0.911	0.093	0.077	19.221
851- 900	8.	881.5	7052.	1.299	1.557	0.762	53.224
901- 950	6.	929.0	5574.	1.027	0.235	0.172	19.494
951-1000	1.	988.0	988.	0.182	0.013	0.012	34.496
1001-1050	3.	1026.7	3080.	0.567	0.366	0.069	70.000
1051-1100	0.	0.0	0.	0.0	0.0	0.0	0.0
1101-1150	1.	1114.0	1114.	0.205	0.221	0.465	0.0
1151-1200	2.	1177.5	2355.	0.434	0.568	0.895	0.0
1201-1250	1.	1240.0	1240.	0.228	0.479	0.771	95.741
1251-1300	2.	1286.5	2573.	0.474	0.255	0.355	5.763
1301-1350	2.	1327.5	2655.	0.489	0.054	0.035	1.887
1351-1400	0.	0.0	0.	0.0	0.0	0.0	0.0
1401-1450	1.	1420.0	1420.	0.262	0.123	0.045	70.000
1451-1500	7.	1470.3	10292.	1.895	1.911	3.330	82.645
1501-1550	1.	1545.0	1545.	0.285	0.010	0.003	1.691
1551-1600	2.	1586.0	3172.	0.584	0.042	0.022	2.726
1601-1650	1.	1640.0	1640.	0.302	0.008	0.003	0.0
1651-1700	10.	1662.0	16620.	3.061	0.589	0.398	1.986
1701-1750	2.	1702.5	3405.	0.627	0.235	0.140	70.000
1751-1800	1.	1790.0	1790.	0.330	0.603	0.193	69.911
1801-1850	0.	0.0	0.	0.0	0.0	0.0	0.0
1851-1900	2.	1870.0	3740.	0.689	0.490	0.400	33.960
1901-1950	1.	1931.0	1931.	0.356	0.002	0.001	0.0
1951-2000	0.	0.0	0.	0.0	0.0	0.0	0.0
2001-2050	0.	0.0	0.	0.0	0.0	0.0	0.0
2051-2100	1.	2100.0	2100.	0.387	0.046	0.098	56.513
2101-2150	0.	0.0	0.	0.0	0.0	0.0	0.0
2151-2200	0.	0.0	0.	0.0	0.0	0.0	0.0
2201-2250	0.	0.0	0.	0.0	0.0	0.0	0.0
2251-2300	0.	0.0	0.	0.0	0.0	0.0	0.0
2301-2350	0.	0.0	0.	0.0	0.0	0.0	0.0
2351-2400	0.	0.0	0.	0.0	0.0	0.0	0.0
2401-2450	0.	0.0	0.	0.0	0.0	0.0	0.0
2451-6000	8.	4397.9	35183.	6.479	9.834	16.998	60.845
	4960.	119.4	542999.	100.000	100.000 2268.16 MM-MMBTU/Y	100.000 1761325.98 TONS/YEAR	43.466

TABLE 3.9

INDUSTRIAL BOILERS, COAL AND OIL FIRED

BOILER STATISTICS BY STATE

STATE	# OF BOILERS	AVERAGE CAPACITY	TOTAL CAPACITY	% OF U.S. CAPACITY	% OF TOTAL FUEL BURNED	% OF TOTAL SULFUR	PERCENT LOAD FACTOR
ALABAMA	55.	530.1	26507.	4.882	1.470	1.343	70.000
ALASKA	0.	0.0	0.	0.0	0.0	0.0	0.0
ARIZONA	3.	1370.3	4111.	0.757	0.021	0.009	0.0
ARKANSAS	4.	349.3	1048.	0.193	0.020	0.003	85.446
CALIFORNIA	10.	27.7	83.	0.015	0.360	0.065	11.945
COLORADO	17.	0.0	0.	0.0	0.016	0.062	0.0
CONNECTICUT	109.	77.5	8215.	1.513	0.949	0.262	40.785
DELAWARE	54.	107.3	5687.	1.047	1.025	0.597	0.0
D. C.	7.	939.6	4698.	0.865	0.416	0.154	64.774
FLORIDA	16.	10.2	163.	0.030	0.040	0.028	38.715
GEORGIA	111.	128.3	13980.	2.575	2.056	1.425	27.618
HAWAII	75.	51.0	2038.	0.375	0.027	0.505	25.633
IDAHO	10.	252.7	2274.	0.419	0.189	0.088	0.0
ILLINOIS	395.	94.8	35156.	6.474	5.619	6.030	40.569
INDIANA	318.	159.2	49180.	9.057	8.178	12.155	37.020
IOWA	0.	0.0	0.	0.0	0.0	0.0	0.0
KANSAS	23.	233.9	5145.	0.948	0.100	0.064	0.0
KENTUCKY	100.	131.4	12216.	2.250	1.880	2.508	37.647
LOUISIANA	3.	210.3	631.	0.116	0.016	0.007	67.977
MAINE	134.	81.9	10896.	2.007	2.956	2.449	70.000
MARYLAND	214.	78.8	16710.	3.077	0.539	1.089	48.748
MASSACHUSETT	393.	56.9	22347.	4.115	2.892	2.158	13.734
MICHIGAN	342.	120.0	38388.	7.070	10.924	10.331	46.263
MINNESOTA	114.	172.1	18582.	3.422	5.037	4.788	24.251
MISSISSIPPI	0.	0.0	0.	0.0	0.0	0.0	0.0
MISSOURI	41.	161.0	6440.	1.186	0.593	0.880	36.249
MONTANA	10.	80.6	725.	0.134	0.161	0.110	51.589
NEBRASKA	11.	146.7	1614.	0.297	0.153	0.081	52.891
NEVADA	3.	305.3	916.	0.169	0.258	0.070	84.998
NEW HAMPSHIR	72.	69.3	4709.	0.867	0.872	0.619	0.0
NEW JERSEY	326.	80.6	26253.	4.835	4.699	1.447	44.509
NEW MEXICO	1.	15.0	15.	0.003	0.002	0.002	0.0
NEW YORK	0.	0.0	0.	0.0	0.0	0.0	0.0
N. CAROLINA	0.	0.0	0.	0.0	0.0	0.0	0.0
NORTH DAKOTA	8.	142.4	997.	0.184	0.141	0.068	39.509
OHIO	469.	142.6	65323.	12.030	11.936	17.335	48.064
OKLAHOMA	5.	123.2	616.	0.113	0.004	0.002	12.921
OREGON	37.	0.0	0.	0.0	0.523	0.0	0.0
PENNSYLVANIA	499.	158.9	78807.	14.513	21.764	22.964	53.220
RHODE ISLAND	61.	59.5	3627.	0.668	0.504	0.365	0.0
S. CAROLINA	67.	281.1	11243.	2.071	2.281	1.685	0.0
SOUTH DAKOTA	4.	43.3	173.	0.032	0.025	0.015	43.225
TENNESSEE	126.	137.8	16817.	3.097	3.282	1.677	3.494
TEXAS	0.	0.0	0.	0.0	0.0	0.0	0.0
UTAH	20.	930.0	930.	0.171	0.0	0.0	0.0
VERMONT	36.	22.8	799.	0.147	0.115	0.078	0.0
VIRGINIA	198.	96.7	17800.	3.278	4.686	2.964	45.106
WASHINGTON	134.	88.3	9975.	1.837	0.791	0.646	34.159
W. VIRGINIA	117.	0.0	0.	0.0	0.0	0.0	0.0
WISCONSIN	199.	83.4	16170.	2.978	2.303	2.774	36.097
WYOMING	9.	124.4	995.	0.183	0.178	0.096	35.934
	4960.	119.4	542999.	100.000	100.000	100.000	43.466
					2268.16 MM-MMBTU/Y	1761325.98 TONS/YEAR	

TABLE 3.10

U.S. ACID PLANT STATISTICS

BY PLANT SIZE AND PLANT TYPE

STATISTICS BY PLANT SIZE

SIZE (100 PCT ACID EQUIV. / DAY)	NO.	AVG. PLANT SIZE (TONS / DAY)	AVG. PLANT AGE (YEARS)	AVG. LOAD FACTOR (PCT)	SULFUR EMITTED / YEAR (PCT. OF TOTAL U.S.) ACID MIST	SO2	TOTAL
0- 100	49.	54.8	26.9	94.50	0.31	2.42	2.73
101- 200	40.	124.6	20.8	93.34	0.80	3.66	4.46
201- 300	31.	217.7	20.0	95.28	1.31	5.01	6.32
301- 400	26.	321.7	19.3	95.18	1.57	6.07	7.64
401- 500	27.	418.5	18.6	95.23	2.04	8.46	10.50
501- 600	12.	508.3	21.8	96.56	1.59	4.59	6.18
601- 700	17.	613.2	17.3	95.22	1.93	7.83	9.75
801- 900	16.	762.5	15.7	95.03	2.50	9.62	12.12
901-1000	10.	905.0	15.5	94.01	1.25	6.00	7.25
1001-2000	17.	1371.8	10.2	94.05	4.47	15.64	20.11
2001-3000	4.	2000.0	11.3	95.90	1.28	5.78	7.07
3001-4000	1.	3050.0	7.0	95.90	0.33	1.96	2.29
4001-5000	1.	4800.0	5.0	95.90	0.51	3.08	3.60
	251.	442.3	19.8	94.73	19.88	80.12	100.00

STATISTICS BY PLANT TYPE

PLANT TYPE	NO.	AVG. PLANT SIZE (TONS / DAY)	AVG. PLANT AGE (YEARS)	AVG. LOAD FACTOR (PCT)	SULFUR EMITTED / YEAR (PCT. OF TOTAL U.S.) ACID MIST	SO2	TOTAL
-1-	37.	69.6	30.9	94.26	0.27	2.71	2.98
-2-	95.	348.1	22.7	94.98	5.65	21.92	27.57
-3-	46.	860.5	7.7	94.71	4.47	25.15	29.62
-4-	49.	457.7	22.2	94.31	6.03	22.01	28.04
-5-	20.	598.8	6.4	95.38	3.46	8.32	11.78
	251.	442.3	19.8	94.73	19.88	80.12	100.00

TYPE1=CHAMBER PLANT TYPE2=SULFUR BURNING WITH 3 CONVERTERS TYPE3=SULFUR BURNING WITH 4 CONVERTERS
TYPE4=WET GAS CONTACT PLANT WITH 3 CONVERTERS TYPE5=WET GAS CONTACT PLANT WITH 4 CONVERTERS

*** THE TOTAL U.S. SULFUR EMISSION FROM ACID PLANTS IS 0.653805E 06 TONS PER YEAR ***

TABLE 3.11

U.S. ACID PLANTS
STATISTICS BY STATE

STATE	NO.	AVG. PLANT SIZE (TONS / DAY)	AVG. PLANT AGE (YEARS)	AVG. LOAD FACTOR (PCT)	SULFUR EMITTED / YEAR (PCT. OF TOTAL U.S.)		
					ACID MIST	SO2	TOTAL
ALABAMA	9.	107.2	21.6	96.17	0.20	0.68	0.88
ALASKA	0.	0.0	0.0	0.0	0.0	0.0	0.0
ARIZONA	5.	357.0	8.8	93.80	0.43	1.62	2.06
ARKANSAS	3.	408.3	18.3	95.90	0.21	0.87	1.08
6 CALIFORNIA	16.	389.1	17.3	94.31	1.13	4.69	5.82
COLORADO	2.	162.5	15.5	92.64	0.06	0.25	0.31
CONNECTICUT	0.	0.0	0.0	0.0	0.0	0.0	0.0
DELAWARE	1.	1000.0	23.0	95.90	0.43	0.96	1.39
1 FLORIDA	31.	868.5	13.8	93.50	2.82	16.98	19.79
GEORGIA	16.	122.5	28.4	94.91	0.21	1.54	1.75
HAWAII	1.	115.0	8.0	95.30	0.02	0.07	0.09
IDAHO	4.	612.5	11.8	95.83	0.29	1.65	1.95
5 ILLINOIS	17.	433.1	20.6	94.82	1.50	4.96	6.46
INDIANA	4.	493.8	27.3	93.03	0.82	1.85	2.67
IOWA	4.	447.5	20.3	96.65	0.22	1.18	1.40
KANSAS	2.	350.0	24.5	95.41	0.14	0.59	0.73
KENTUCKY	2.	362.5	24.0	96.17	0.19	0.47	0.66
3 LOUISIANA	9.	1322.2	14.4	94.96	2.29	7.74	10.02
MAINE	0.	0.0	0.0	0.0	0.0	0.0	0.0
MARYLAND	8.	271.3	27.0	95.26	0.47	1.69	2.16
MASSACHUSETT	1.	300.0	4.0	95.90	0.08	0.19	0.27
MICHIGAN	4.	217.5	24.3	96.39	0.29	0.75	1.04
MINNESOTA	1.	330.0	14.0	95.48	0.04	0.21	0.25
MISSISSIPPI	2.	400.0	19.0	96.81	0.09	0.53	0.62
MISSOURI	6.	308.3	14.2	95.95	0.26	1.19	1.45
MONTANA	1.	450.0	23.0	94.38	0.07	0.43	0.50
NEBRASKA	0.	0.0	0.0	0.0	0.0	0.0	0.0
NEVADA	1.	400.0	20.0	89.05	0.06	0.36	0.42
NEW HAMPSHIR	0.	0.0	0.0	0.0	0.0	0.0	0.0
4 NEW JERSEY	14.	464.3	22.0	95.50	1.72	4.85	6.57
NEW MEXICO	2.	275.0	13.0	96.36	0.07	0.35	0.42
NEW YORK	2.	285.0	33.0	96.52	0.15	0.37	0.53
N. CAROLINA	6.	616.7	15.2	94.26	0.39	2.39	2.78
NORTH DAKOTA	0.	0.0	0.0	0.0	0.0	0.0	0.0
OHIO	10.	217.5	20.9	94.68	0.29	1.38	1.67
OKLAHOMA	2.	280.0	27.5	90.19	0.07	0.51	0.58
OREGON	0.	0.0	0.0	0.0	0.0	0.0	0.0
PENNSYLVANIA	9.	290.0	19.6	96.24	0.68	2.35	3.03
RHODE ISLAND	1.	50.0	24.0	93.16	0.01	0.03	0.04
S. CAROLINA	8.	57.5	28.1	93.54	0.05	0.44	0.49
SOUTH DAKOTA	0.	0.0	0.0	0.0	0.0	0.0	0.0
TENNESSEE	4.	812.5	34.8	97.44	0.59	3.02	3.61
2 TEXAS	19.	576.1	15.2	95.23	2.74	8.46	11.20
UTAH	4.	437.5	14.8	90.88	0.36	2.19	2.55
VERMONT	0.	0.0	0.0	0.0	0.0	0.0	0.0
VIRGINIA	12.	160.8	27.5	94.27	0.26	1.35	1.61
WASHINGTON	3.	118.3	15.3	91.64	0.05	0.32	0.37
W. VIRGINIA	1.	400.0	25.0	92.47	0.04	0.25	0.29
WISCONSIN	2.	150.0	23.0	95.90	0.08	0.19	0.27
WYOMING	2.	162.5	13.0	96.17	0.03	0.21	0.24
C. C.	0.	0.0	0.0	0.0	0.0	0.0	0.0
	251.	442.3	19.8	94.73	19.88	80.12	100.00

TABLE 3.12

U.S. SULFUR PLANTS

STATISTICS BY PLANT SIZE

DAILY SHORT TON CAPACITIES	NUMBER OF PLANTS	AVFRAGE SIZE	AVERAGF AGE	PERCENT OF US. DAILY CAPACITY	PERCENT OF US. ANNUAL SULFUR
0- 100	120.	36.54	8.40	24.846	25.075
101- 200	19.	138.00	9.95	14.857	14.809
201- 300	7.	235.68	5.57	9.348	9.327
301- 400	8.	329.42	3.38	14.933	14.883
401- 500	5.	436.58	9.00	12.369	12.329
501- 600	2.	504.00	6.00	5.712	5.695
601- 700	1.	645.12	9.00	3.656	3.646
701- 800	0.	0.0	0.0	0.0	0.0
801- 900	0.	0.0	0.0	0.0	0.0
901-1000	0.	0.0	0.0	0.0	0.0
1001-1100	0.	0.0	0.0	0.0	0.0
1101-1200	1.	1120.00	4.00	6.346	6.322
1201-1300	0.	0.0	0.0	0.0	0.0
1301-1400	1.	1400.00	1.00	7.933	7.914
TOTALS	164.	107.61	8.13	100.000	100.000
				TOTAL CAPACITY 17647.504	TOTAL SULFUR 437399.000

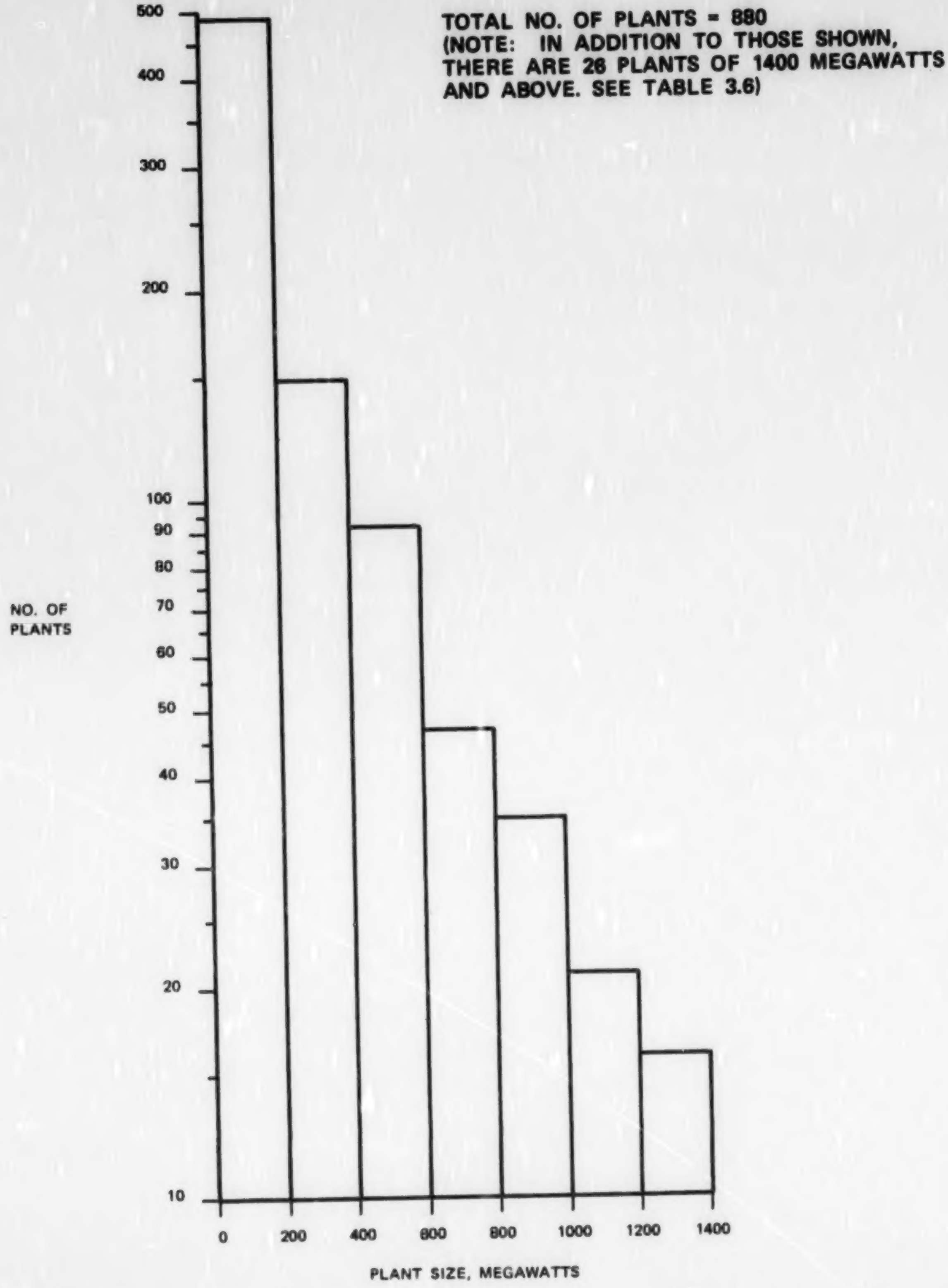
TABLE 3.13

U.S. SULFUR PLANTS - STATISTICS BY STATE

STATES	NUMBER OF PLANTS	AVERAGE SIZE	AVERAGE AGE	REFINERY FEED		NATURAL GAS FEED		BOTH FEEDS	
				PERCENT OF US CAPACITY	PERCENT OF US SULFUR	PERCENT OF US CAPACITY	PERCENT OF US SULFUR	PERCENT OF US CAPACITY	PERCENT OF US SULFUR
ALABAMA	2.	216.2	1.0	2.450	2.446	0.0	0.0	2.450	2.446
ALASKA	1.	10.1	1.0	0.0	0.0	0.057	0.057	0.057	0.057
ARIZONA	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ARKANSAS	4.	51.8	13.8	1.015	1.008	0.159	0.158	1.174	1.166
CALIFORNIA	18.	167.8	8.5	0.0	0.0	17.110	17.055	17.110	17.055
COLORADO	1.	20.2	5.0	0.0	0.0	0.114	0.114	0.114	0.114
CONNECTICUT	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
DELAWARE	2.	434.0	9.0	0.0	0.0	4.919	4.904	4.919	4.904
D. C.	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
FLORICA	4.	185.9	1.0	4.214	4.195	0.0	0.0	4.214	4.195
GEORGIA	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HAWAII	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
IDAHO	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
ILLINOIS	4.	159.3	4.8	0.0	0.0	3.611	3.601	3.611	3.601
INDIANA	3.	154.6	1.3	0.0	0.0	2.627	2.618	2.627	2.618
IOWA	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
KANSAS	2.	24.6	5.0	0.0	0.0	0.279	0.274	0.279	0.274
KENTUCKY	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
LOUISIANA	6.	106.4	6.8	0.0	0.0	3.617	3.601	3.617	3.601
MAINE	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MARYLAND	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MASSACHUSETT	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MICHIGAN	3.	33.2	11.0	0.0	0.0	0.565	0.560	0.565	0.560
MINNESOTA	2.	95.2	7.5	0.0	0.0	1.079	1.075	1.079	1.075
MISSISSIPPI	4.	371.6	5.8	8.231	8.212	0.190	0.190	8.422	8.402
MISSOURI	1.	89.6	2.0	0.0	0.0	0.508	0.503	0.508	0.503
MONTANA	3.	87.0	4.7	0.0	0.0	1.479	1.475	1.479	1.475
NEBRASKA	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NEVADA	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NEW HAMPSHIF	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NEW JERSEY	7.	103.5	7.0	0.0	0.0	4.106	4.092	4.106	4.092
NEW MEXICO	7.	23.5	9.1	0.743	0.737	0.190	0.189	0.933	0.926
NEW YORK	1.	56.0	4.0	0.0	0.0	0.317	0.320	0.317	0.320
N. CAROLINA	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NORTH DAKOTA	2.	136.1	9.0	0.127	0.127	1.415	1.416	1.542	1.543
OHIO	3.	35.5	5.3	0.038	0.038	0.565	0.568	0.603	0.606
OKLAHOMA	2.	12.9	9.0	0.146	0.149	0.0	0.0	0.146	0.149
OREGON	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
PENNSYLVANIA	6.	74.7	8.3	0.698	0.695	1.840	2.162	2.539	2.858
RHODE ISLAND	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
S. CAROLINA	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SOUTH DAKOTA	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TENNESSEE	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TEXAS	58.	94.0	8.8	23.622	23.537	7.286	7.258	30.907	30.795
UTAH	2.	12.3	3.5	0.063	0.062	0.076	0.075	0.140	0.137
VERMONT	0.	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
VIRGINIA	1.	56.0	16.0	0.0	0.0	0.317	0.320	0.317	0.320
WASHINGTON	1.	22.4	11.0	0.127	0.126	0.0	0.0	0.127	0.126
W. VIRGINIA	1.	30.2	13.0	0.171	0.171	0.0	0.0	0.171	0.171
WISCONSIN	1.	16.8	1.0	0.0	0.0	0.095	0.091	0.095	0.091
WYOMING	12.	85.8	13.2	5.832	5.818	0.0	0.0	5.832	5.818
TOTALS ----	164.	107.6	8.1	47.477	47.323	52.523	52.677	100.000	100.000
EQUIVALENTS-				8378.578	206990.500	9268.973	230408.500	17647.551	437399.000

FIGURE 3.1

DISTRIBUTION OF U.S. UTILITY PLANTS WITH PLANT SIZE



TOTAL NO. OF PLANTS = 880
(NOTE: IN ADDITION TO THOSE SHOWN, THERE ARE 26 PLANTS OF 1400 MEGAWATTS AND ABOVE. SEE TABLE 3.6)

FIGURE 3.2

SIZE DISTRIBUTION OF U.S. UTILITY BOILERS

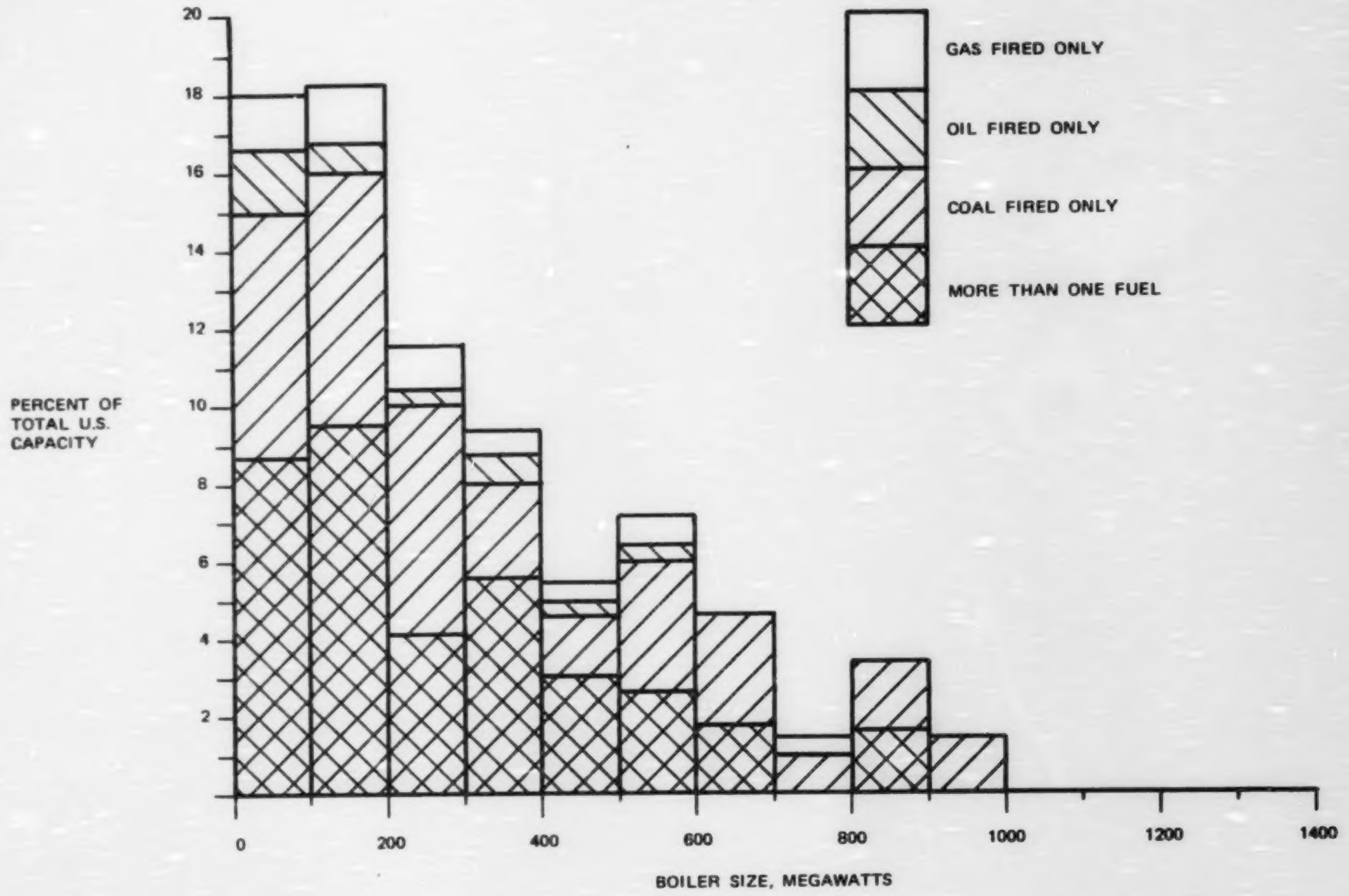


FIGURE 3.3

DISTRIBUTION OF PLANT LOAD FACTORS FOR THE U.S. UTILITY INDUSTRY

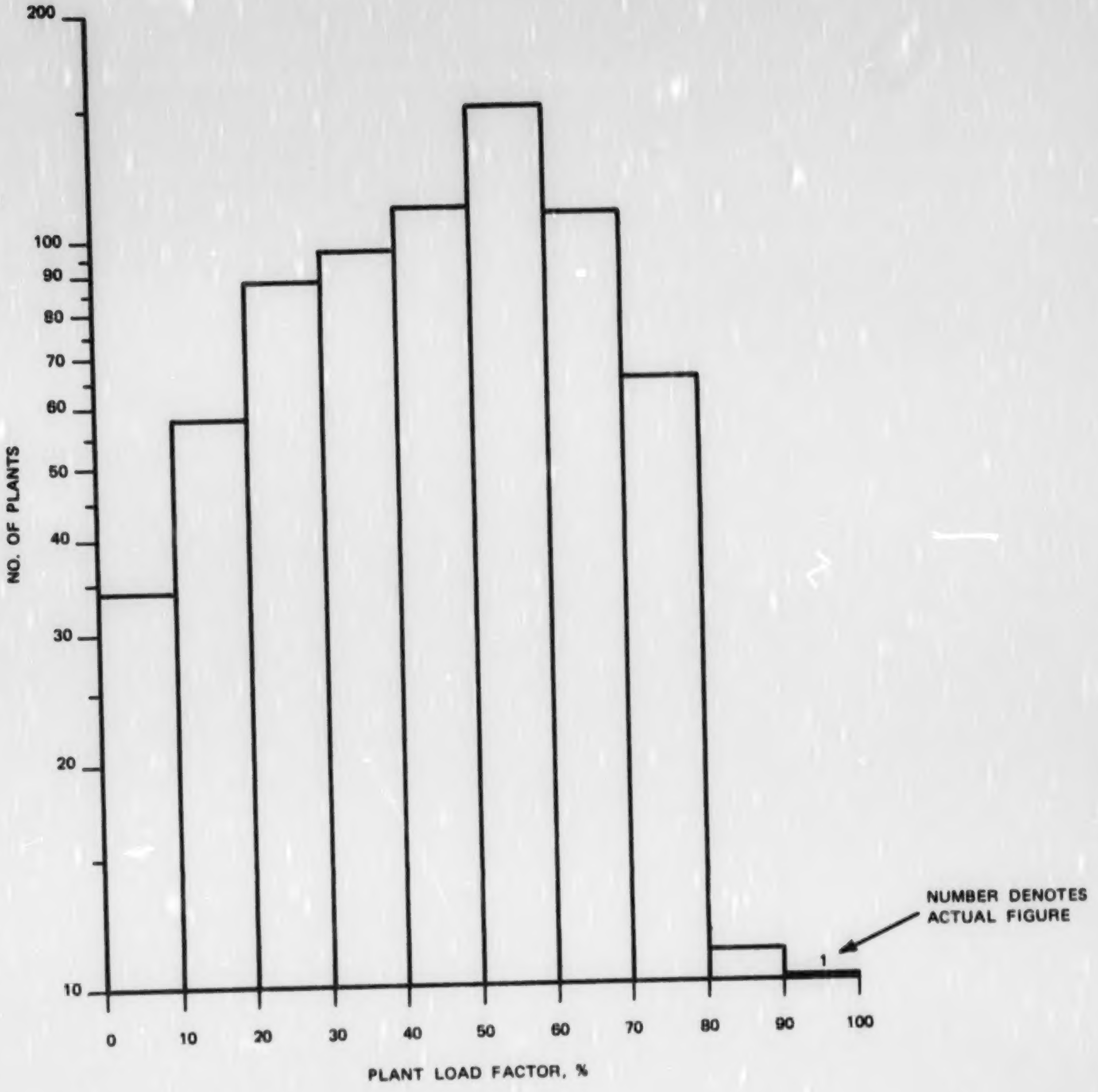


FIGURE 3.4

VARIATION OF PLANT LOAD FACTOR WITH UTILITY PLANT SIZE

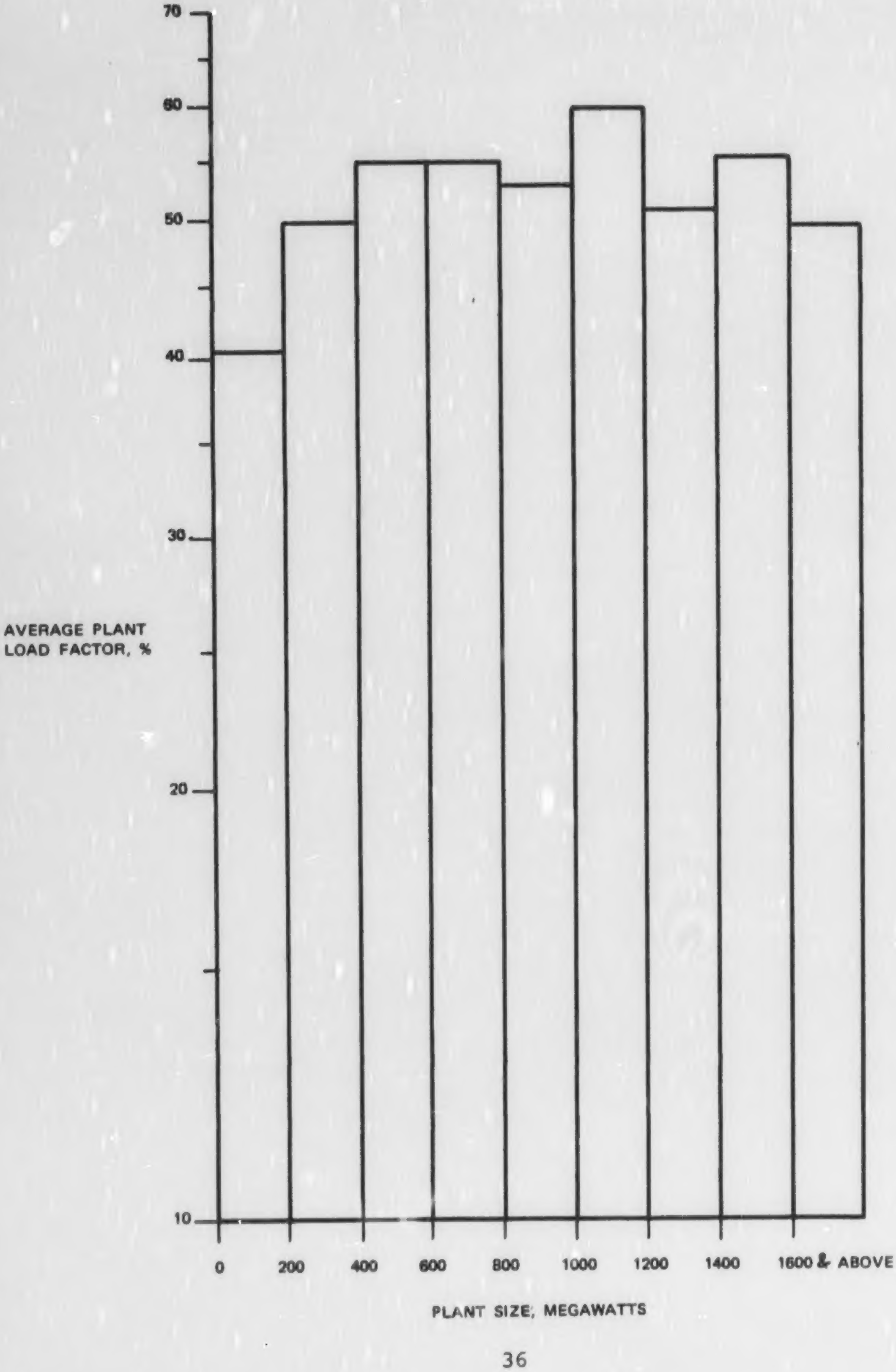
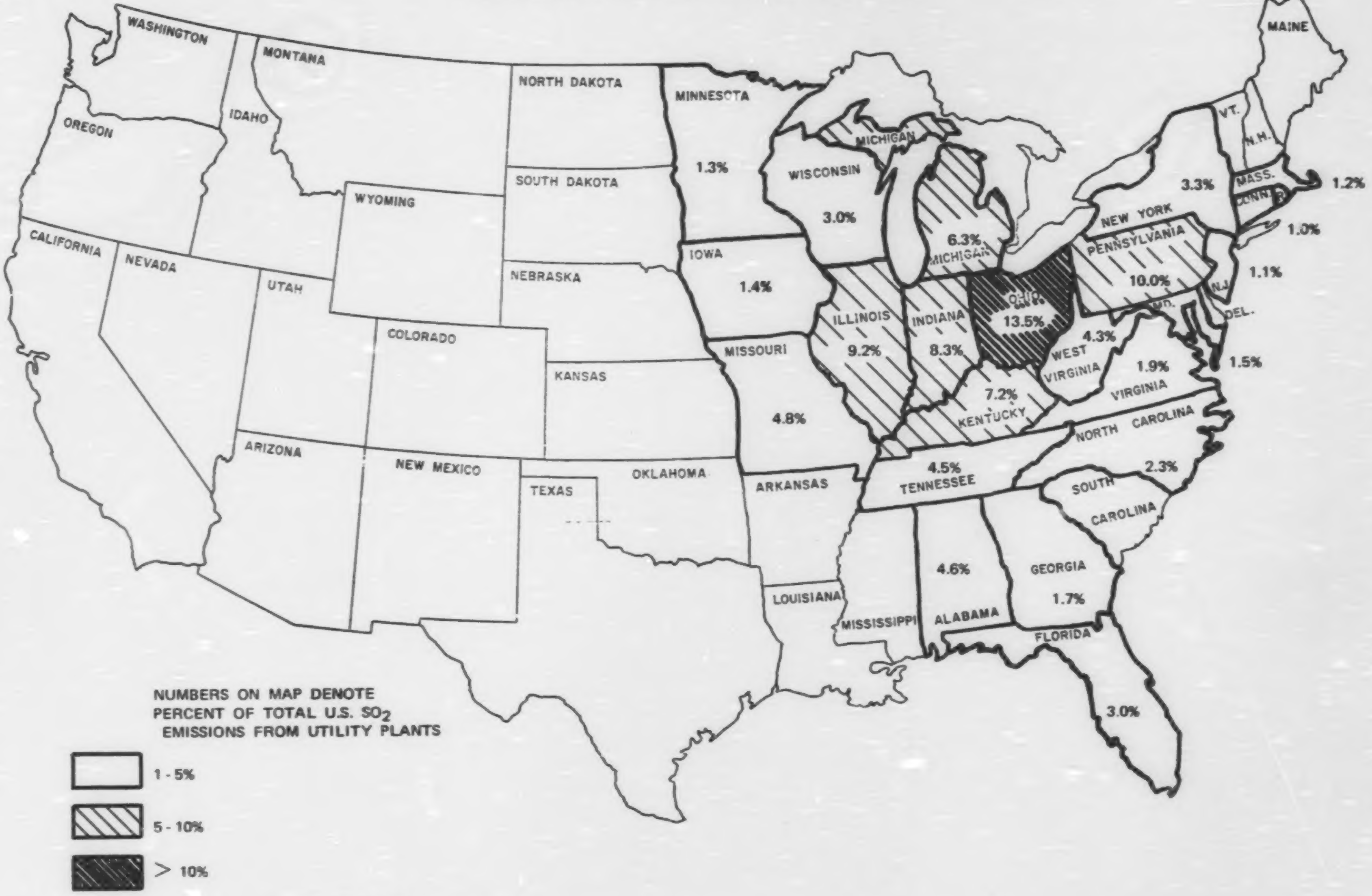


FIGURE 3.5

GEOGRAPHICAL DISTRIBUTION OF SO_2 EMISSIONS FROM THE UTILITY INDUSTRY


WASHINGTON
MONTANA
NORTH DAKOTA
MINNESOTA
1.3%
MAINE
VT.
N.H.
OREGON
IDAHO
MICHIGAN
WISCONSIN
3.0%
SOUTH DAKOTA
WYOMING
3.3%
MASS.
CONN.
R.I.
1.2%
NEW YORK
1.0%
CALIFORNIA
NEVADA
IOWA
1.4%
6.3%
MICHIGAN
PENNSYLVANIA
10.0%
N.J.
1.1%
UTAH
NEBRASKA
ILLINOIS
9.2%
INDIANA
8.3%
OHIO
13.5%
MD.
DEL.
COLORADO
KANSAS
MISSOURI
4.8%
WEST VIRGINIA
4.3%
VIRGINIA
1.9%
1.5%
7.2%
KENTUCKY
NORTH CAROLINA
2.3%
ARIZONA
NEW MEXICO
TEXAS
OKLAHOMA
ARKANSAS
4.5%
TENNESSEE
SOUTH CAROLINA
4.6%
GEORGIA
1.7%
LOUISIANA
MISSISSIPPI
ALABAMA
FLORIDA
3.0%

NUMBERS ON MAP DENOTE PERCENT OF TOTAL U.S. SO_2 EMISSIONS FROM UTILITY PLANTS

- ☐ 1 - 5%
- ▨ 5 - 10%
- ■ > 10%

FIGURE 3.6

SO_2 EMISSIONS FROM U.S. UTILITY PLANTS

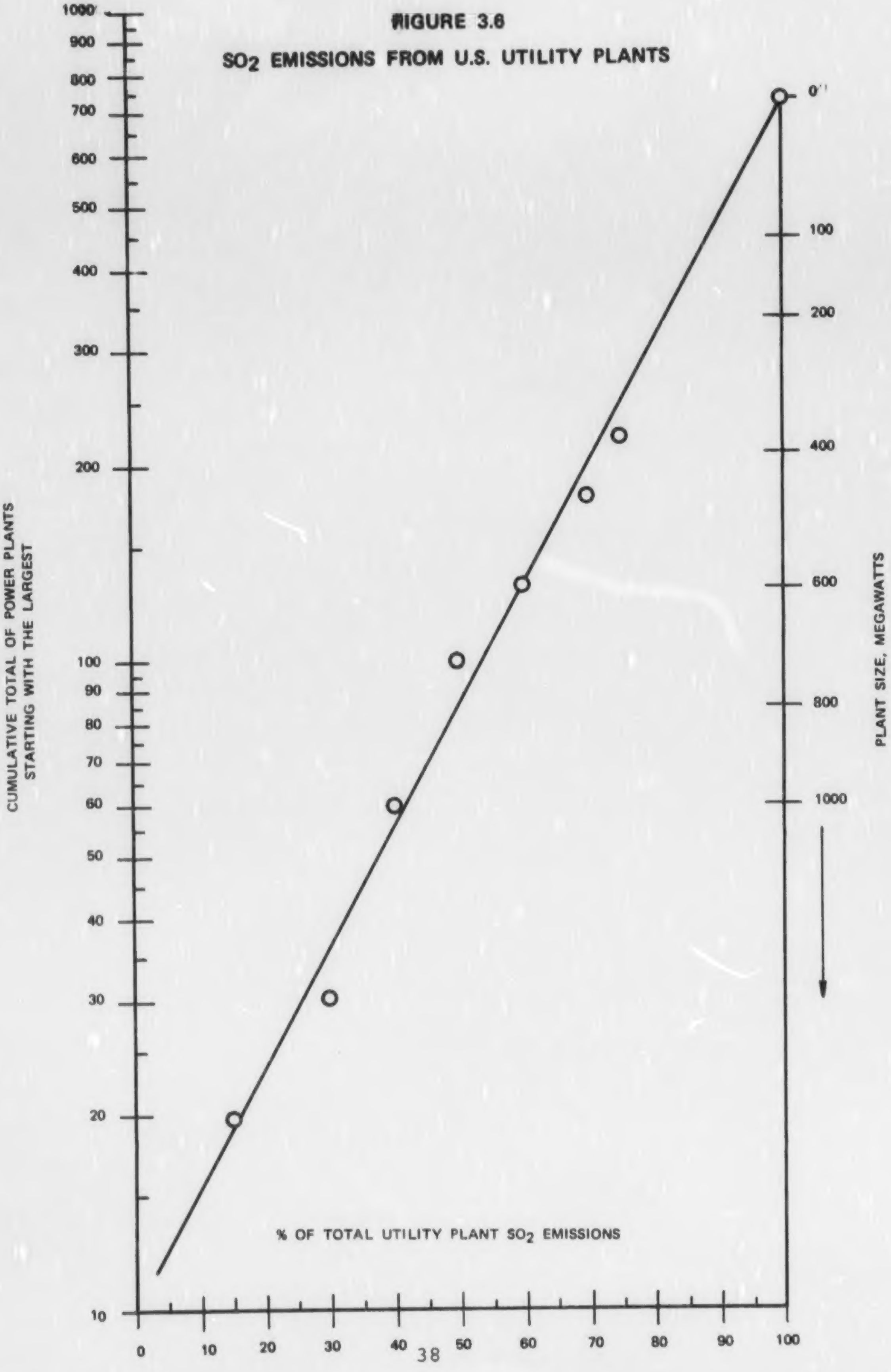


FIGURE 3.7

DISTRIBUTION OF BOILERS AND AVERAGE PLANT AGE WITH UTILITY PLANT SIZE

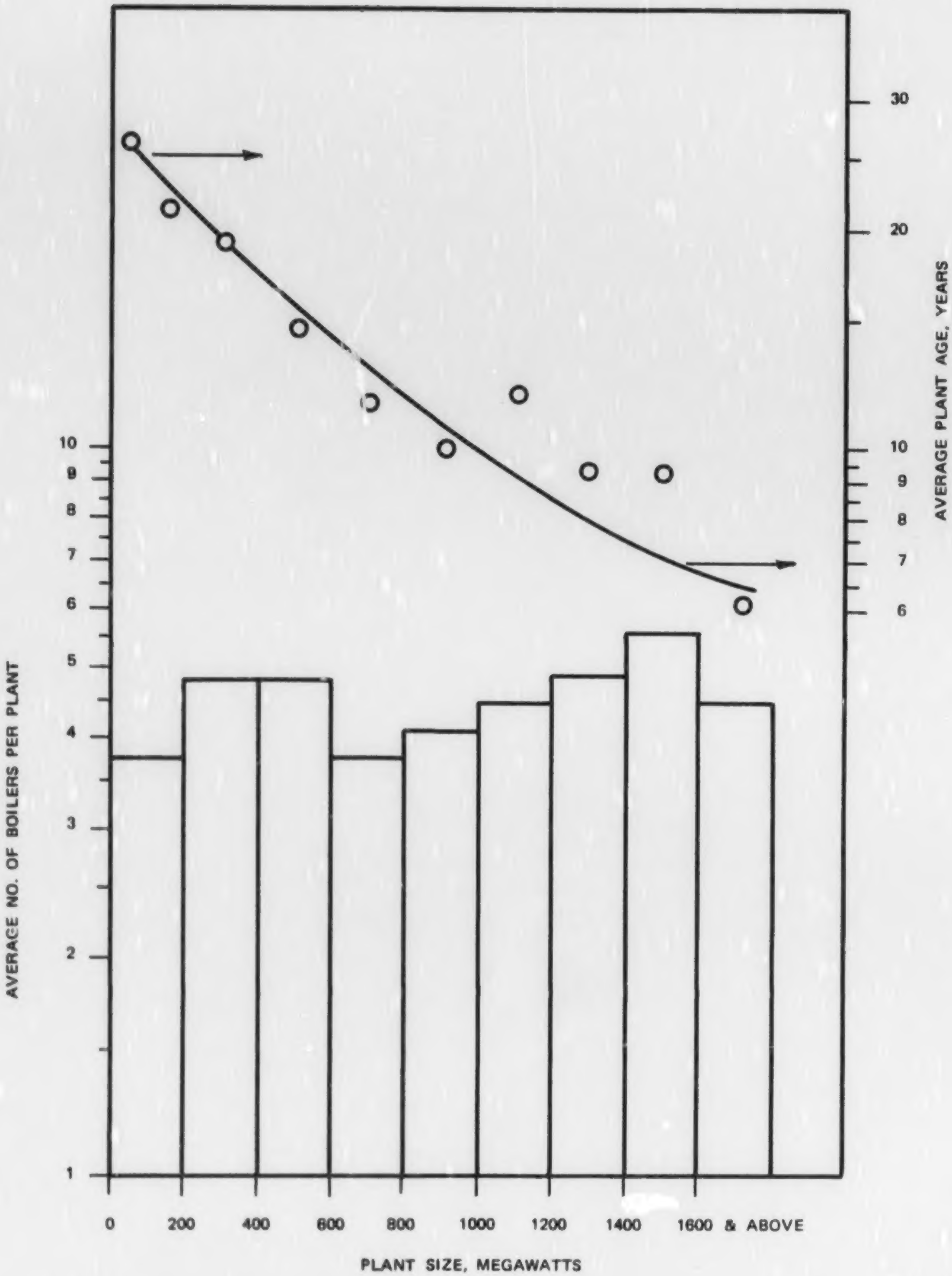


FIGURE 3.8

DISTRIBUTION OF BOILERS AND BOILER AGE WITH UTILITY BOILER SIZE

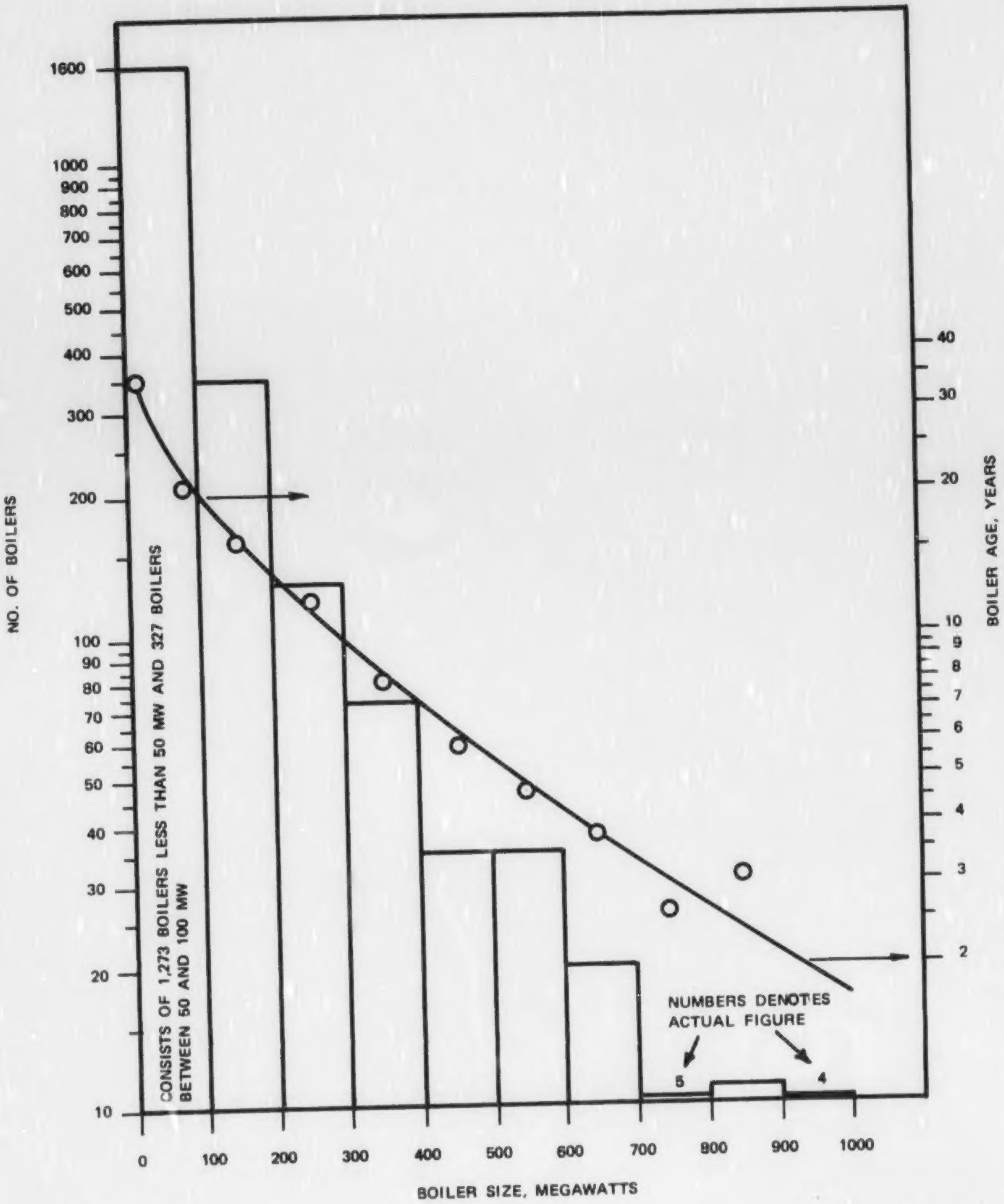


FIGURE 3.9

DISTRIBUTION OF U.S. SMELTERS WITH PLANT SIZE

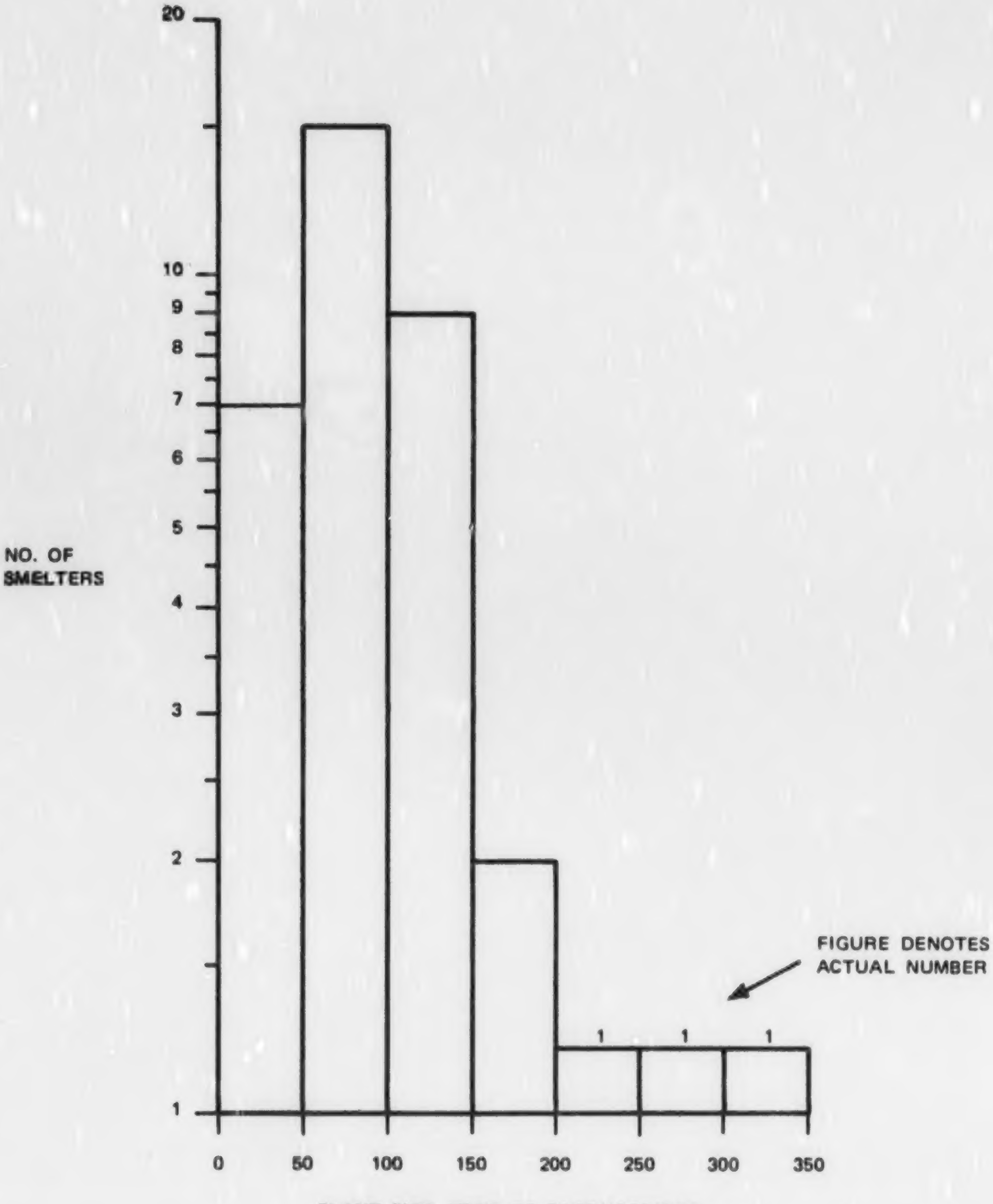


FIGURE 3.10

SO_2 EMISSIONS FROM U.S. SMELTERS

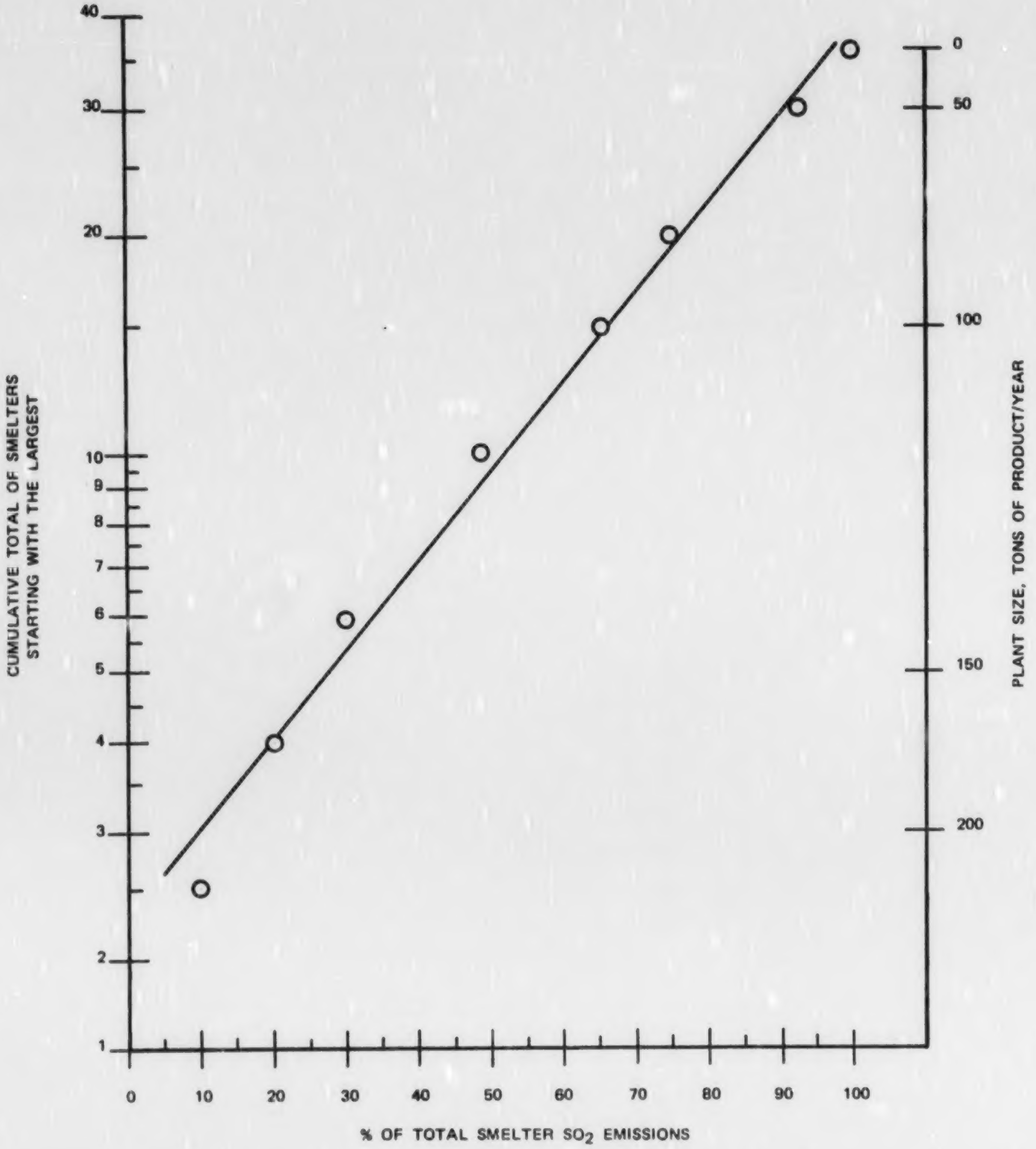
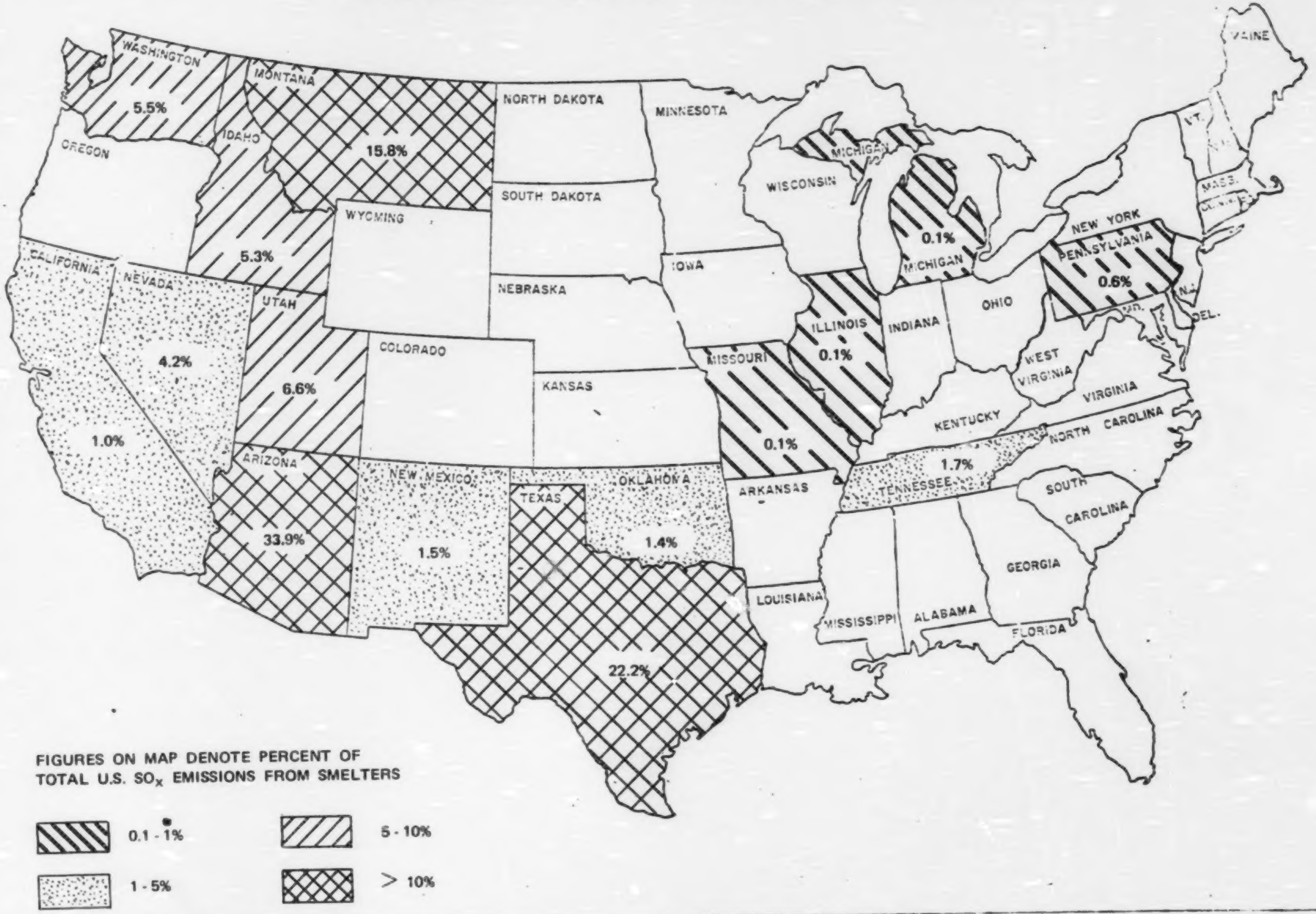


FIGURE 3.11

GEOGRAPHICAL DISTRIBUTION OF SO_2 EMISSIONS FROM SMELTERS


WASHINGTON
5.5%
OREGON
CALIFORNIA
1.0%
NEVADA
4.2%
IDAHO
5.3%
MONTANA
15.8%
UTAH
6.6%
ARIZONA
33.9%
WYOMING
COLORADO
NEW MEXICO
1.5%
NORTH DAKOTA
SOUTH DAKOTA
NEBRASKA
KANSAS
OKLAHOMA
1.4%
TEXAS
22.2%
MINNESOTA
IOWA
MISSOURI
0.1%
ARKANSAS
LOUISIANA
WISCONSIN
MICHIGAN
0.1%
MICHIGAN
ILLINOIS
0.1%
INDIANA
OHIO
KENTUCKY
TENNESSEE
1.7%
MISSISSIPPI
ALABAMA
GEORGIA
FLORIDA
SOUTH CAROLINA
NORTH CAROLINA
VIRGINIA
WEST VIRGINIA
NEW YORK
PENNSYLVANIA
0.6%
MD.
N.J.
DEL.
VT.
MASS.
MAINE
0.1 - 1%
5 - 10%
1 - 5%
> 10%

FIGURES ON MAP DENOTE PERCENT OF TOTAL U.S. SO_x EMISSIONS FROM SMELTERS

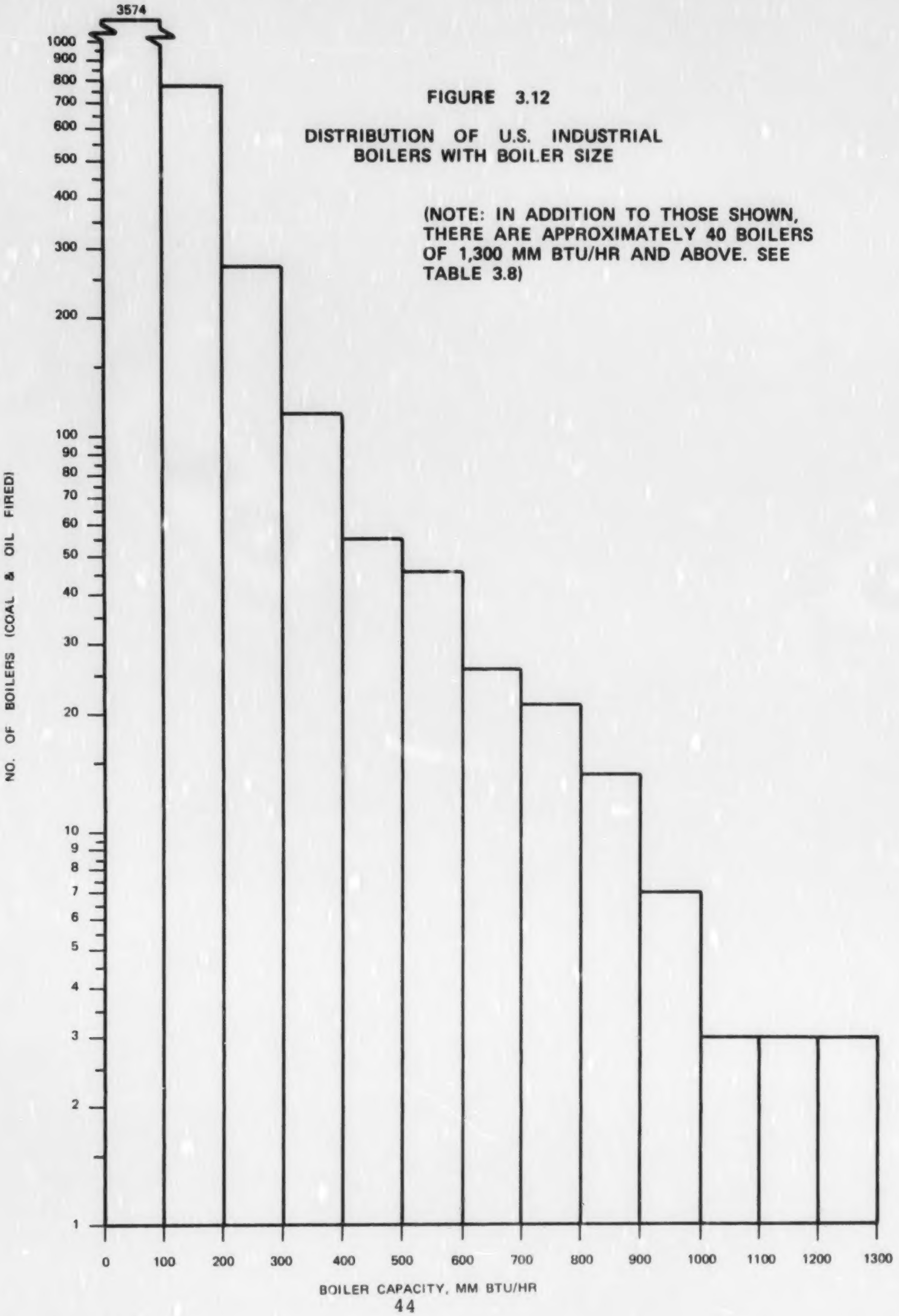


FIGURE 3.12

DISTRIBUTION OF U.S. INDUSTRIAL BOILERS WITH BOILER SIZE

(NOTE: IN ADDITION TO THOSE SHOWN, THERE ARE APPROXIMATELY 40 BOILERS OF 1,300 MM BTU/HR AND ABOVE. SEE TABLE 3.8)

FIGURE 3.13

SIZE DISTRIBUTION OF U.S. INDUSTRIAL BOILERS

(NOTE: IN ADDITION TO THOSE SHOWN, THERE ARE 40 BOILERS IN THE RANGE OF 1000 - 6000 MM BTU/HR WHICH MAKE UP 17.6% OF TOTAL U.S. CAPACITY. SEE TABLE 3.8.)

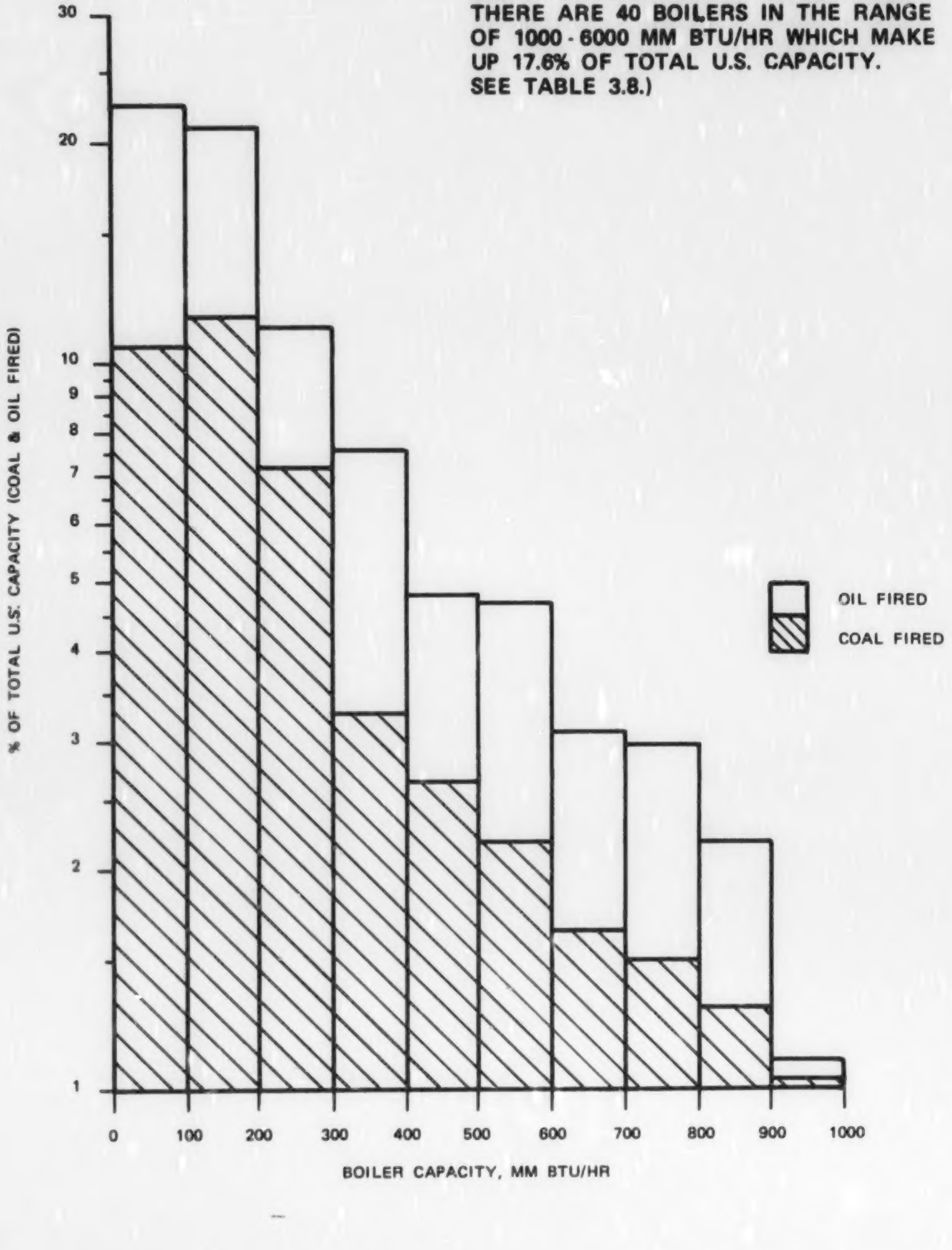


FIGURE 3.14

GEOGRAPHICAL DISTRIBUTION OF SO_2 EMISSIONS FROM INDUSTRIAL BOILERS

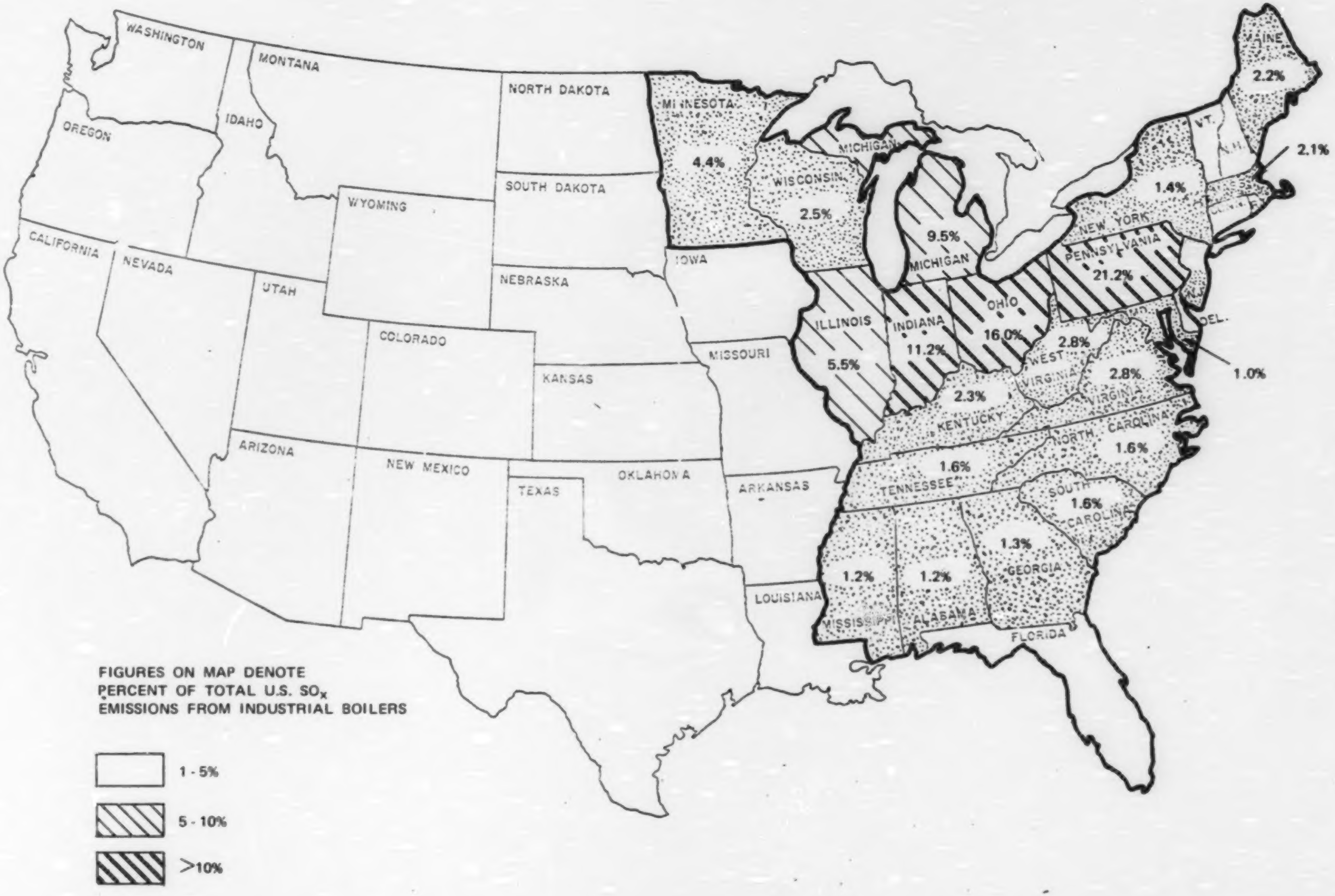

WASHINGTON
MONTANA
NORTH DAKOTA
OREGON
IDAHO
MINNESOTA
4.4%
WISCONSIN
2.5%
MICHIGAN
9.5%
MICHIGAN
MAINE
2.2%
VT.
N.H.
2.1%
1.4%
NEW YORK
PENNSYLVANIA
21.2%
SOUTH DAKOTA
WYOMING
CALIFORNIA
NEVADA
UTAH
NEBRASKA
IOWA
ILLINOIS
5.5%
INDIANA
11.2%
OHIO
16.0%
DEL.
1.0%
WEST VIRGINIA
2.8%
VIRGINIA
2.8%
COLORADO
MISSOURI
KANSAS
KENTUCKY
2.3%
NORTH CAROLINA
1.6%
ARIZONA
NEW MEXICO
TEXAS
OKLAHOMA
ARKANSAS
TENNESSEE
1.6%
SOUTH CAROLINA
1.6%
GEORGIA
1.3%
LOUISIANA
MISSISSIPPI
1.2%
ALABAMA
1.2%
FLORIDA
FIGURES ON MAP DENOTE
PERCENT OF TOTAL U.S. SO_x
EMISSIONS FROM INDUSTRIAL BOILERS
1 - 5%
5 - 10%
>10%

FIGURE 3.15

SO_2 EMISSIONS FROM U.S. INDUSTRIAL BOILERS

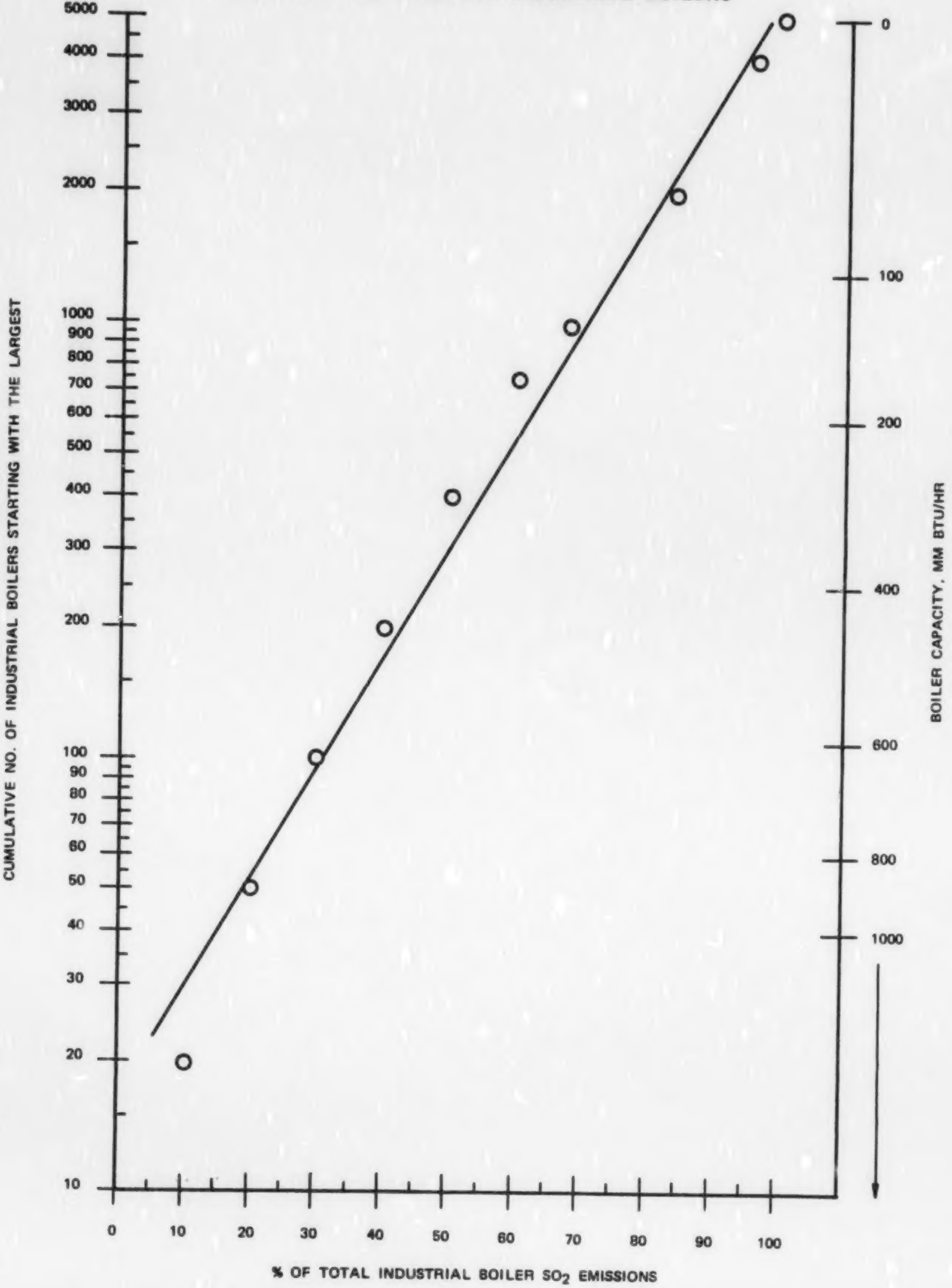


FIGURE 3.16

DISTRIBUTION OF U.S. SULFURIC ACID PLANTS WITH PLANT SIZE

(NOTE: IN ADDITION TO THOSE SHOWN, THERE ARE 23 PLANTS IN THE RANGE OF 1000 - 5000 TONS/DAY OF 100% ACID EQUIVALENT. SEE TABLE 3.10.)

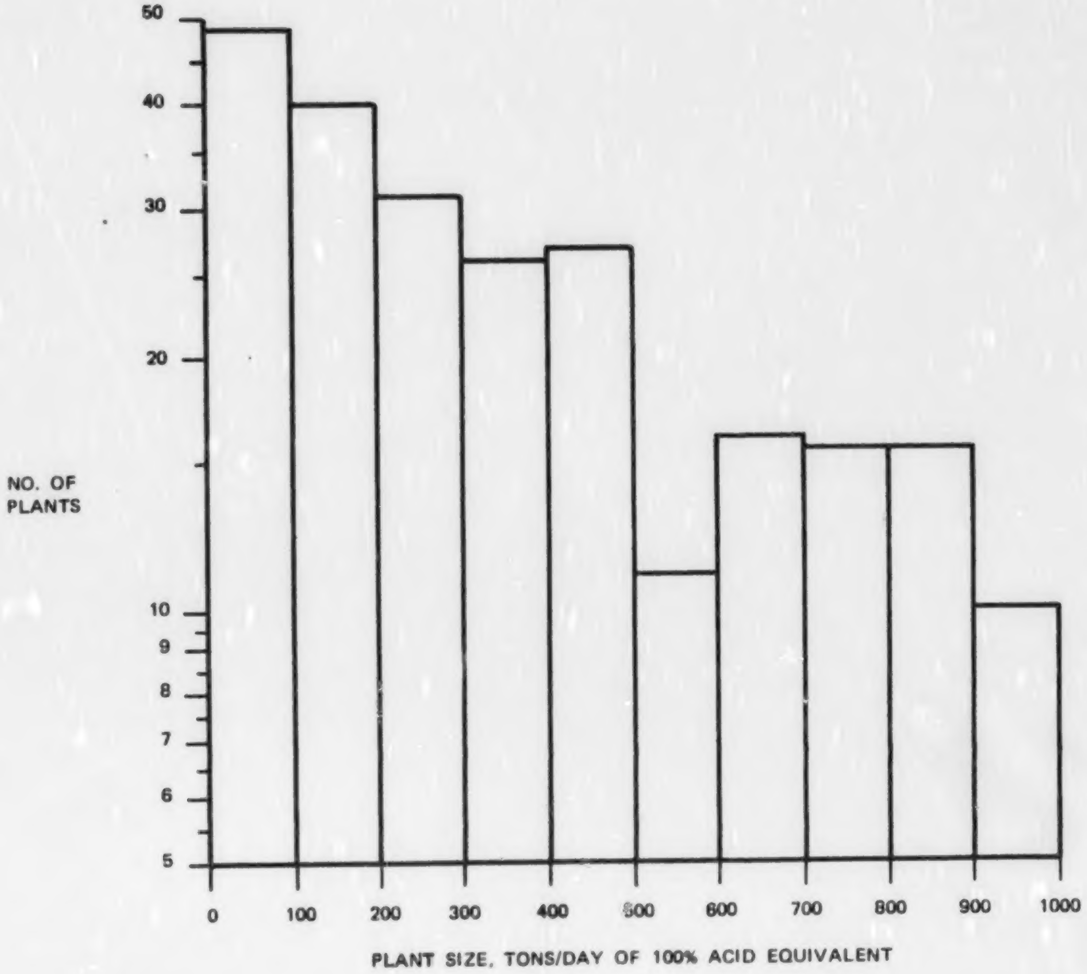
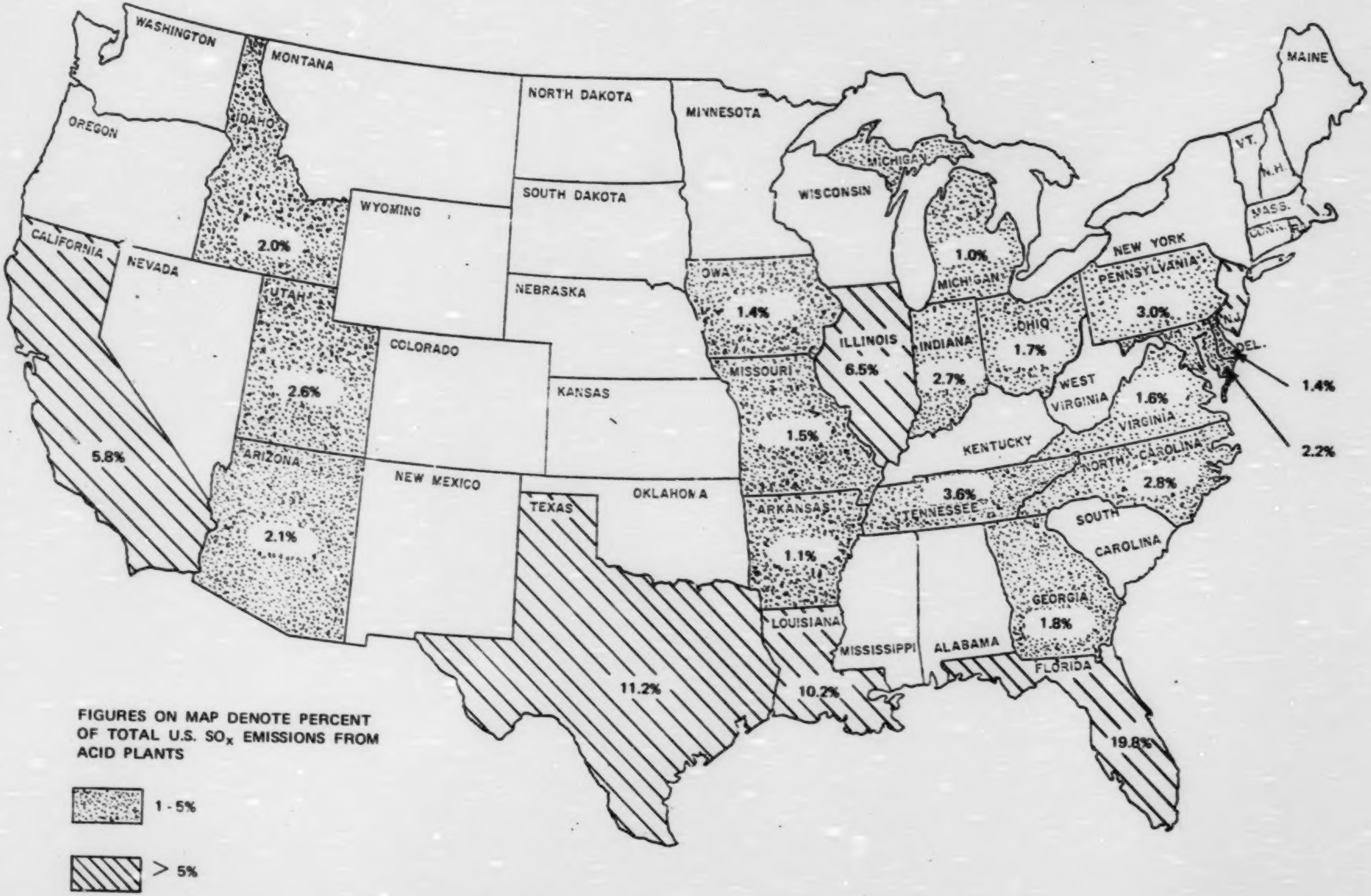


FIGURE 3.17

GEOGRAPHICAL DISTRIBUTION OF SO_2 EMISSIONS FROM SULFURIC ACID PLANTS


WASHINGTON
OREGON
CALIFORNIA
5.8%
NEVADA
IDAHO
2.0%
MONTANA
WYOMING
UTAH
2.6%
ARIZONA
2.1%
COLORADO
NEW MEXICO
NORTH DAKOTA
SOUTH DAKOTA
NEBRASKA
KANSAS
OKLAHOMA
TEXAS
11.2%
MINNESOTA
IOWA
1.4%
MISSOURI
1.5%
ARKANSAS
1.1%
LOUISIANA
10.2%
WISCONSIN
MICHIGAN
1.0%
MICHIGAN
ILLINOIS
6.5%
INDIANA
2.7%
OHIO
1.7%
KENTUCKY
TENNESSEE
3.6%
MISSISSIPPI
ALABAMA
GEORGIA
1.8%
FLORIDA
19.8%
SOUTH CAROLINA
NORTH CAROLINA
2.8%
VIRGINIA
1.6%
WEST VIRGINIA
PENNSYLVANIA
3.0%
NEW YORK
N.J.
DEL.
1.4%
2.2%
VT.
N.H.
MASS.
CONN.
MAINE
1 - 5%
> 5%

FIGURES ON MAP DENOTE PERCENT OF TOTAL U.S. SO_x EMISSIONS FROM ACID PLANTS

FIGURE 3.18

SO_2 EMISSIONS FROM U.S. SULFURIC ACID PLANTS

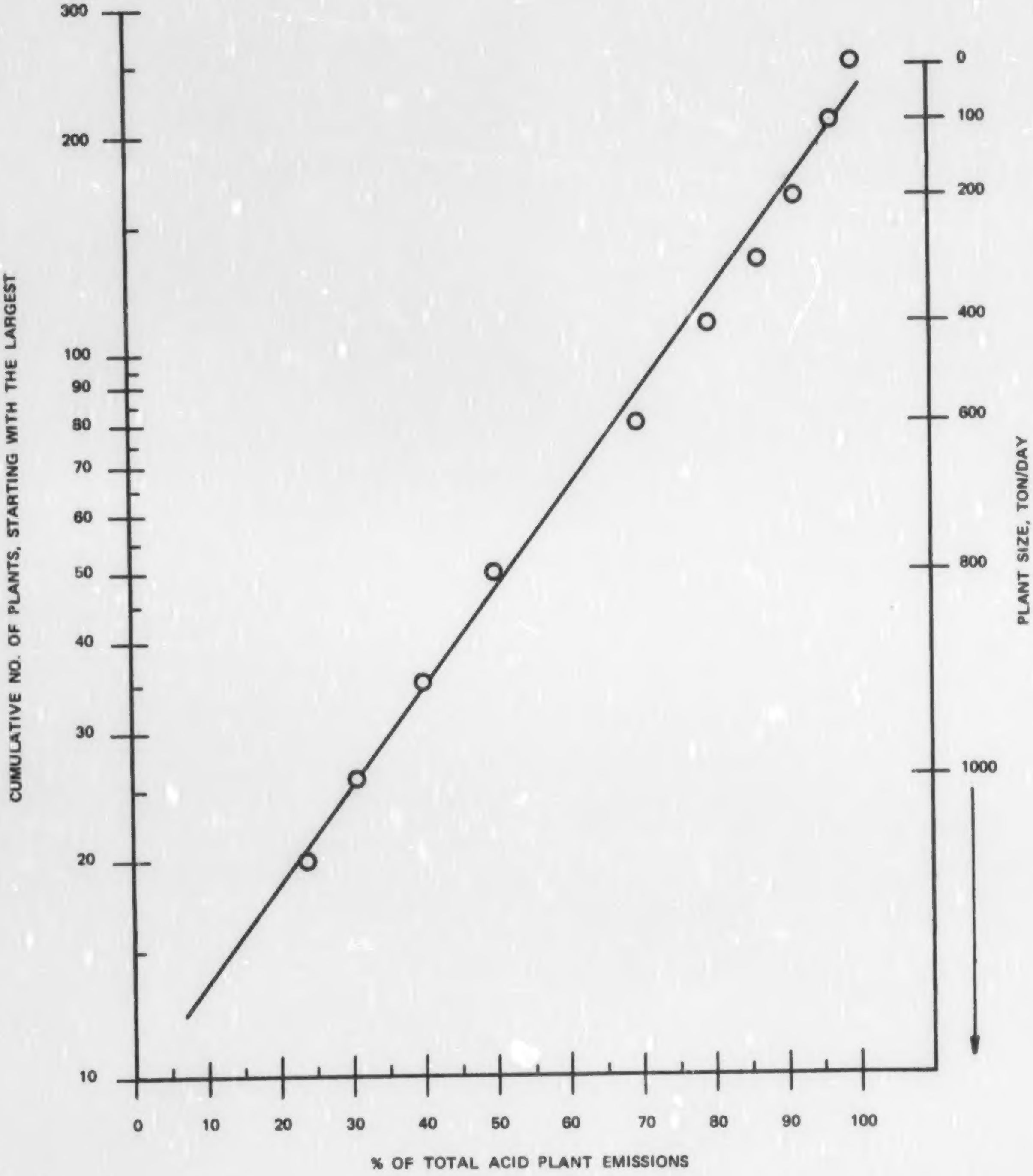


FIGURE 3.19

DISTRIBUTION OF U.S. SULFUR PLANTS WITH PLANT SIZE

(NOTE: IN ADDITION TO THOSE SHOWN, THERE ARE 2 PLANTS WITH SIZE GREATER THAN 100 TONS/DAY. SEE TABLE 3.12.)

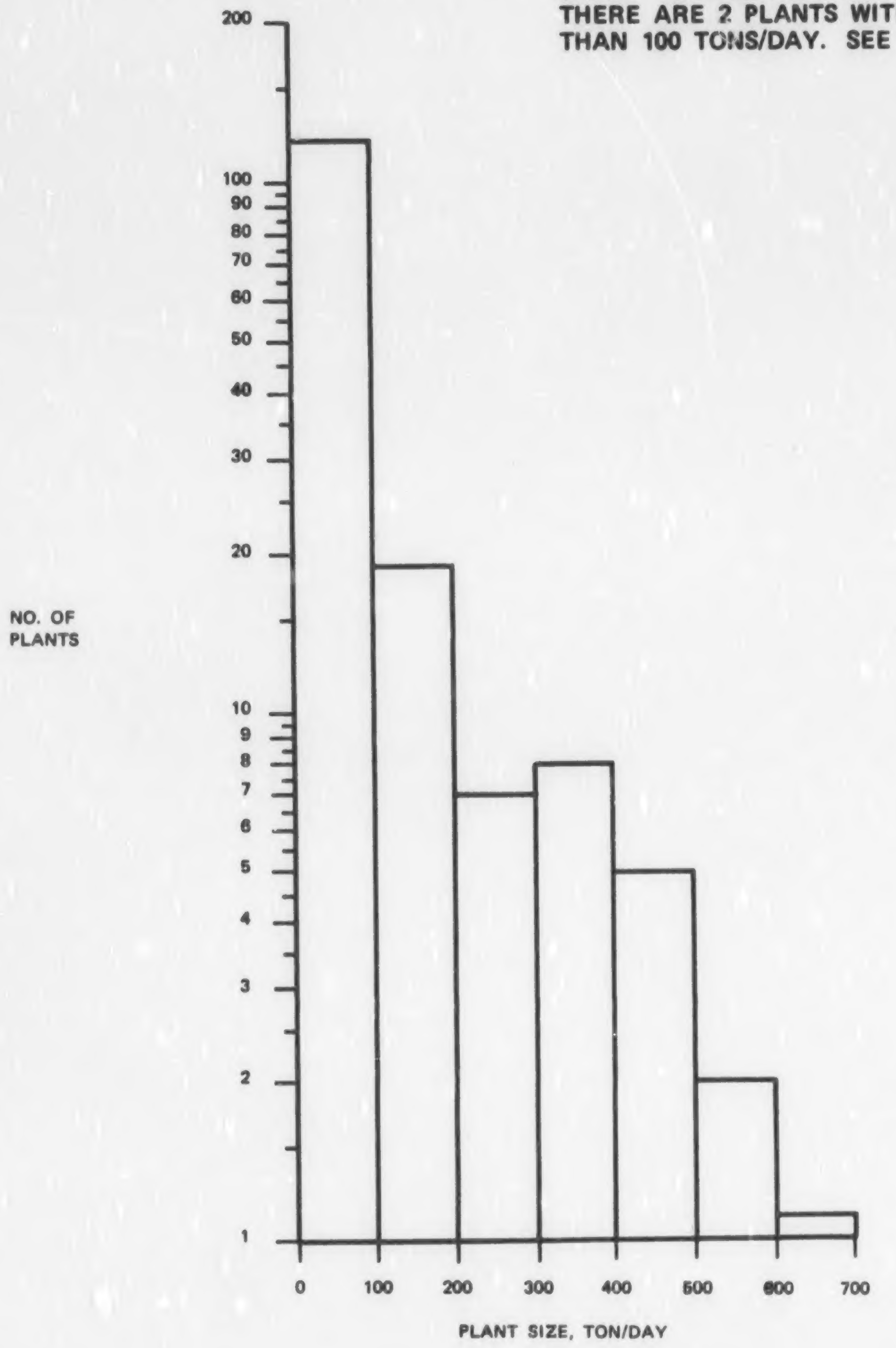
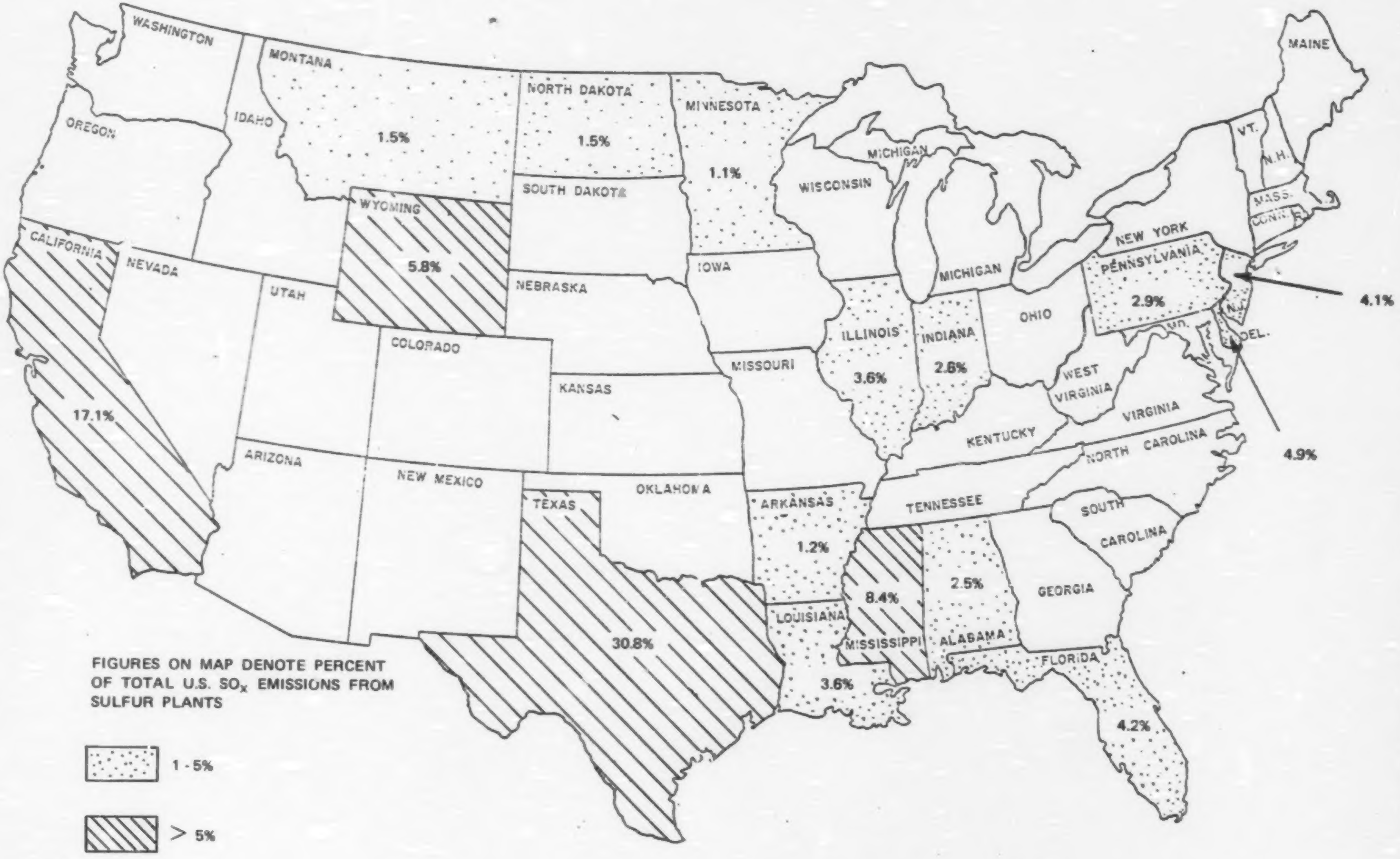


FIGURE 3.20

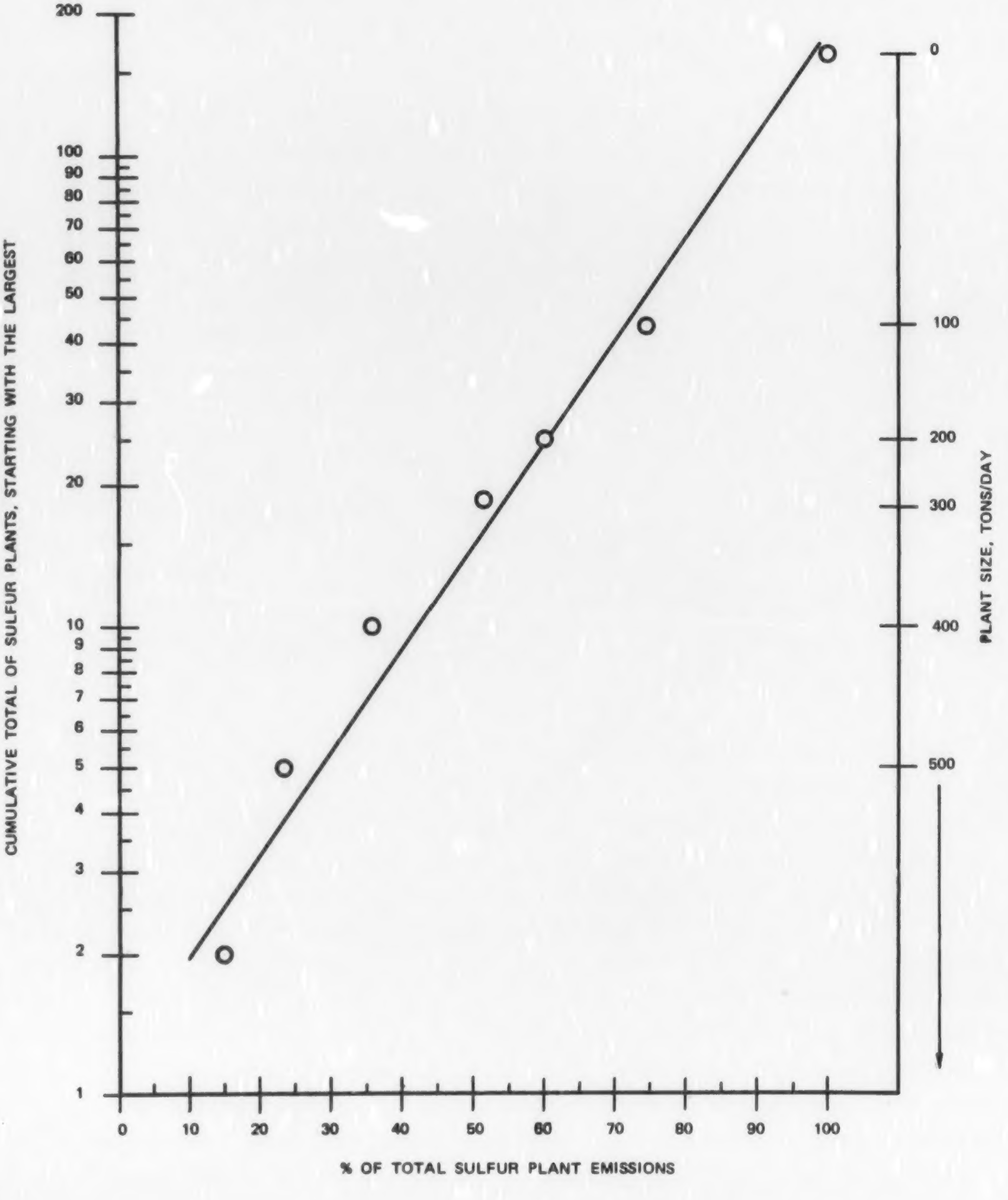
GEOGRAPHICAL DISTRIBUTION OF SO_2 EMISSIONS FROM SULFUR PLANTS


WASHINGTON
OREGON
CALIFORNIA
17.1%
NEVADA
IDAHO
MONTANA
1.5%
WYOMING
5.8%
UTAH
ARIZONA
COLORADO
NEW MEXICO
NORTH DAKOTA
1.5%
SOUTH DAKOTA
NEBRASKA
KANSAS
OKLAHOMA
TEXAS
30.8%
MINNESOTA
1.1%
IOWA
MISSOURI
ARKANSAS
1.2%
LOUISIANA
3.6%
WISCONSIN
MICHIGAN
MICHIGAN
ILLINOIS
3.6%
INDIANA
2.6%
OHIO
KENTUCKY
TENNESSEE
MISSISSIPPI
8.4%
ALABAMA
2.5%
GEORGIA
FLORIDA
4.2%
SOUTH CAROLINA
NORTH CAROLINA
VIRGINIA
WEST VIRGINIA
MD.
DEL.
4.9%
N.J.
4.1%
PENNSYLVANIA
2.9%
NEW YORK
VT.
N.H.
MASS.
CONN.
R.
MAINE
1 - 5%
> 5%

FIGURES ON MAP DENOTE PERCENT OF TOTAL U.S. SO_x EMISSIONS FROM SULFUR PLANTS

FIGURE 3.21

SO_2 EMISSIONS FROM U.S. SULFUR PLANTS



4. THE GENERAL MODEL

4.1 The General Process Model

The plants in the models have, as far as possible, been made self-contained apart from the intake of basic raw feed materials; i.e., the plant should not be buying natural gas or electricity. If possible, it should not even be buying desulfurized fuel oil since supply cannot be assumed. There are obviously exceptions if the plant is an addition to a larger conventional plant; e.g., with stack gas scrubbing for a power plant it would be illogical not to assume a supply of power. In general, a large plant having a coal feed will generate its own power, steam and heat requirements by burning coal and scrubbing the stack gases.

It was not a primary concern to provide special chemical by-products from any process, but to avoid additional treatment facilities for impure materials by routing these side streams back to the plant fuel supply where possible. This approach simplifies the models and minimizes the effect of credits for special chemical by-products on the plant costs.

The cost of equipment and raw material, utility and waste product quantities have all been related to one or more basic process parameters; e.g., in the stack gas scrubbing models, the basic process parameters are flue gas flow rate and sulfur content of the fuel. For a plant producing high quality fuel, the basic process parameters are product flow rate and properties of the raw feed materials.

Where possible, equipment costs were related directly to the basic process parameters. However, the format of some of the estimates used to develop the models prevented this. In these cases, the available cost information was carefully examined relative to the General Cost Model to determine exactly what the costs included.

The equipment costs were extracted from these estimates by using the relationships between construction labor costs, other material costs and equipment costs given in the General Cost Model.

Each plant design was examined to fix maximum train sizes for each group of equipment. It has been assumed that N trains cost N times the cost of one train. Where a plant is largely made up of several trains, size variations were only taken in increments of their size.

For the smaller plants, it was possible to examine the cost of every item of equipment and assign an exponent of size to give cost variations. However, for the larger plants, whole sections have been grouped together. The following is given as a general guide to the exponents for equipment cost vs. size (9,14,21):

	n, $\left[\frac{\text{Cost}_2}{\text{Cost}_1} = \left(\frac{\text{Size}_2}{\text{Size}_1}\right)^n\right]$
Increasing number of trains of equipment	1.0
Blowers	0.9
Solids grinding equipment	0.8
Steam generation equipment	0.8
Process furnaces and reformers	0.7
Compressors	0.7
Power generation equipment	0.7
Solids handling equipment	0.6
Offsites	0.6
Other process units	0.6

4.2 The General Cost Model

4.2.1 Bases For Costs

All costs in the models are those in existence at the end of 1973. To update prior cost information used in the construction of the models, an annual inflation multiplication factor of 1.05 has been used. All costs other than unit costs for labor, raw materials, etc., are shown in thousands of dollars (M$).

The direct field construction labor cost, L, and the direct cost of operating labor, CO, both refer to a Gulf Coast (Houston) location. For any other location, they are adjusted through the use of a location factor, F, which is explained in section 4.3.

Whenever possible in the development of the cost models discussed in this report, major equipment costs, E, have been related to plant size variations. The reference values of E have been taken from actual plant cost estimates when these were available. Sometimes, however, the cost estimates were not available in such a detailed breakdown. In such cases, the relationships developed in the General Cost Model were used to analyze the cost data. The relationships in the General Cost Model were developed based on procedures reported and recommended in the literature (9,13) and on Kellogg's general experience.

4.2.2 Capital Cost Model

Major equipment costs, E, represent the cost of major equipment delivered to the site, but not located, tied-in to piping, instruments, etc., or commissioned. It includes material costs only. Major equipment is defined to include furnaces, heat exchangers, converters, reactors, towers, drums and tanks, pumps, compressors, transportation and conveying equipment, special equipment (filters, centrifuges, dryers, agitators, grinding equipment, cyclones, etc.), and major gas ductwork.

Other material costs, M, represent the cost of piping, electrical, process instrumentation, paint, insulation, foundations, concrete structures, and structural steel for equipment support. It does not include such items as site preparation, steel frame structures, process buildings, cafeterias, control rooms, shops, offices, etc.

M has been taken as a fixed fraction of E. Whenever possible, this fraction has been determined from an estimate covering the particular plant under consideration. This fraction is often different for each section of the plant. If particular details were not available, the following relationships have been assumed (9):

Solids handling plant:	M = 0.40E
Chemical process plant:	M = 0.80E

Direct field construction labor costs, L, are based on Gulf Coast rates and productivities. Again, L has been taken as a fixed fraction of E. Whenever possible, it has been derived from an estimate covering the particular plant under consideration. This fraction is often different for each section of the plant. If particular details were not available, the following relationships have been assumed (9):

Solids handling plant:	L = 0.40E
Chemical process plant:	L = 0.60E

Indirect costs associated with field labor have been assumed as follows:

Fringe benefits and payroll burden	= 0.12 L
Field administration, supervision temporary facilities	= 0.17 L
Construction equipment and tools	= 0.14 L
Total field labor indirect costs	= 0.43 L

Home office engineering includes home office construction, engineering and design, procurement, client services, accounting, cost engineering, travel and living expenses, reproduction and communication. This could range from under 10% to almost 20% of the major equipment and other material costs. In the model, this has been assumed to be 15% of the total direct material cost (E + M).

The bare cost of the plant, BARC, is defined as the sum of equipment costs, other material costs, construction labor and labor indirects, and home office engineering. For a Gulf Coast location, it is given by:

$$\begin{aligned} \text{BARC} &= E + M + L + 0.43\ L + 0.15\ (E + M) \\ &= 1.15\ (E + M) + 1.43\ L \end{aligned}$$

For any other location, it is given by:

$$\text{BARC} = 1.15\ (E + M) + 1.43\ L \cdot F$$

where F is the location factor (see section 4.3).

Taxes and insurance can be 1-4% of the bare cost. In the model, they have been assumed to be 2%. Contractor's overheads and profit could depend on several factors, but are generally in the range of 6-13% of the bare cost. A value of 10% was chosen for the model.

A contingency has been included in the model and is expressed as a fraction of the bare cost. It represents the degree of uncertainty in the process design and the cost estimate. The contingency, CONTIN, could range from zero for a well-established process to 0.20 or more for a process still under development.

The total plant investment, TPI, is defined as the sum of the bare cost (including contingency), taxes and insurance, and contractor's overheads and profit. It is therefore given by:

$$\begin{aligned} \text{TPI} &= (1.0 + \text{CONTIN})\ \text{BARC} + 0.02\ (1.0 + \text{CONTIN})\ \text{BARC} \\ &\quad + 0.10\ (1.0 + \text{CONTIN})\ \text{BARC} \\ &= 1.12\ (1.0 + \text{CONTIN})\ \text{BARC} \end{aligned}$$

In order to obtain the total capital required for construction of a particular plant, some additional costs should be added to the total plant investment. These costs are:

1. Start-up costs
2. Working capital
3. Interest during construction

Start-up costs, STC, have been assumed to be 20% of the total net annual operating cost, AOC (see section 4.2.3 for explanation of AOC). Thus:

$$\text{STC} = 0.20\ \text{AOC}$$

Working capital, WKC, is required for raw materials inventory, plant materials and supplies, etc. For simplification, it has also been assumed to be 20% of the total net annual operating cost, AOC.

Thus:

$$\text{WKC} = 0.20\ \text{AOC}$$

Interest during construction, IDC, obviously increases with the length of the construction period which, to some extent, is a function of the size of the plant. The construction of plants the size of the stack gas scrubbing units is now taking about 2-3 years and projects of the magnitude and complexity of a substitute natural gas plant or a power station are taking 4-5 years. Two different values for the interest during construction have therefore been assumed. The first is intended to be used for stack gas scrubbing units fitted to existing power plants or for constructions well under $100 million:

$$IDC = 0.12\ TPI^{*}$$

The second is for the larger, more complex plants such as substitute natural gas, solvent refined coal, and power plants:

$$IDC = 0.18\ TPI^{*}$$

The total capital required, TCR, is equal to the sum of the total plant investment, start-up costs, working capital, and interest during construction.

Thus:

$$TCR = TPI + STC + WKC + IDC$$

For stack gas scrubbing units, this can be reduced to:

$$\begin{aligned} TCR &= TPI + 0.20\ AOC + 0.20\ AOC + 0.12\ TPI \\ &= 1.12\ TPI + 0.40\ AOC \end{aligned}$$

For the larger plants, this can be reduced to:

$$\begin{aligned} TCR &= TPI + 0.20\ AOC + 0.20\ AOC + 0.18\ TPI \\ &= 1.18\ TPI + 0.40\ AOC \end{aligned}$$

From section 4.2.3, AOC is calculated from:

$$AOC = 0.078\ TPI + 2.0\ TO \cdot CO\ (1.0 + F) + ANR$$

---

*See Appendix A for derivation of equation where TO = total number of shift operators

ANR = Annual cost of raw materials, utilities, and waste disposal, less by-product credits.

Therefore, for stack gas scrubbing units, the equation for the total capital required becomes:

$$\begin{aligned}TCR &= 1.12\ TPI + 0.40\ [0.078\ TPI + 2.0\ TO \cdot CO\ (1.0 + F) + ANR] \\ &= 1.12\ TPI + 0.03\ TPI + 0.8\ TO \cdot CO\ (1.0 + F) + 0.40\ ANR \\ &= 1.15\ TPI + 0.8\ TO \cdot CO\ (1.0 + F) + 0.40\ ANR\end{aligned}$$

For the larger plants, the equation for the total capital required becomes:

$$\begin{aligned}TCR &= 1.18\ TPI + 0.40\ [0.078 TPI + 2.0\ TO \cdot CO\ (1.0 + F) + ANR] \\ &= 1.18\ TPI + 0.03\ TPI + 0.8\ TO \cdot CO\ (1.0 + F) + 0.4\ ANR \\ &= 1.21\ TPI + 0.8\ TO \cdot CO\ (1.0 + F) + 0.4\ ANR\end{aligned}$$

The buildup of costs to determine the total capital required is illustrated in Figure 4.1.

4.2.3 Operating Cost Model

The total net annual operating cost, AOC, is the total cost of operating the plant less the credits from the sale of by-products. It does not include return of capital, payment of interest on capital, income tax on equity returns or depreciation. The total net annual operating cost is made up of the following items:

1. Annual cost of raw materials, utilities, and waste disposal, less by-product credits
2. Annual cost of operating labor and supervision
3. Annual cost of maintenance labor and supervision
4. Annual cost of plant supplies and replacements
5. Annual cost of administration and overheads
6. Annual cost of local taxes and insurance

The annual cost of raw materials, utilities, and waste disposal, less by-product credits, ANR, is clearly a function of the particular process under consideration. It is given by different relationships for each model.

The total number of operators employed on all shifts, TO, is different for each process and is either given as an equation or number for each particular model. It has been assumed that each operator works 40 hours per week for 50 weeks per year (2000 hours per year). If CO is the hourly rate for an operator (Gulf Coast basis), then the annual cost of operating labor is given by:

$$\text{Operating labor (Gulf Coast)} = \frac{TO \cdot 2000 \cdot CO}{1000}$$

$$= 2\ TO \cdot CO \qquad \text{M\$/yr}$$

The annual cost of operating labor for any other location has been assumed to be:

$$\text{Operating labor} = 2\ TO \cdot CO\ (0.5 + 0.5\ F)$$

Supervision was assumed to be 15% of operating labor. Thus, the total cost of operating labor and supervision, AOL, is given by:

$$AOL = 1.15\ [2\ TO \cdot CO\ (0.5 + 0.5\ F)]$$

$$= 2.3\ TO \cdot CO\ (0.5 + 0.5\ F)$$

The annual cost of maintenance labor has been assumed to be 1.5% of the total plant investment. Maintenance supervision is 15% of maintenance labor. Therefore, the total annual cost of maintenance labor and supervision, AML, is:

$$\begin{aligned} AML &= 1.15\ (0.015\ TPI) \\ &= 0.018\ TPI\ \text{(rounded up)} \end{aligned}$$

Plant supplies and replacements include charts, cleaning supplies, miscellaneous chemicals, lubricants, paint, and replacement parts such as gaskets, seals, valves, insulation, welding materials, packing, balls (grinding), vessel lining materials, etc. The annual cost of plant supplies and replacements, APS, has been assumed to be 2% of the total plant investment. Thus:

$$APS = 0.02\ TPI$$

Administration and overheads include salaries and wages for administrators, secretaries, typists, etc., office supplies and equipment, medical and safety services, transportation and communications, lighting, janitorial services, plant protection, payroll overheads, employee benefits, etc. The annual cost of administration and overheads, AOH, has been assumed to be 70% of the annual operator, maintenance labor, and total supervision costs. Thus:

$$\begin{aligned} AOH &= 0.70\ [2.3\ TO \cdot CO\ (0.5 + 0.5F) + 0.018\ TPI] \\ &= 1.7\ TO \cdot CO\ (0.5 + 0.5F) + 0.013\ TPI\ \text{(rounded up)} \end{aligned}$$

Local taxes and insurance include property taxes, fire and liability insurance, special hazards insurance, business interruption insurance, etc. The annual local taxes and insurance, ATI, have been assumed to be 2.7% of the total plant investment. Thus:

$$ATI = 0.027\ TPI$$

The total net annual operating cost, AOC, is therefore given by:

$$
\begin{aligned}
\mathrm{AOC} &= \mathrm{ANR} + \mathrm{AOL} + \mathrm{AML} + \mathrm{APS} + \mathrm{AOH} + \mathrm{ATI} \\
&= \mathrm{ANR} + 2.3\ \mathrm{TO{\cdot}CO}\ (0.5 + 0.5F) + 0.018\ \mathrm{TPI} \\
&\quad + 0.02\ \mathrm{TPI} + 1.7\ \mathrm{TO{\cdot}CO}\ (0.5 + 0.5F) + 0.013\ \mathrm{TPI} \\
&\quad + 0.027\ \mathrm{TPI} \\
&= 0.078\ \mathrm{TPI} + 4.0\ \mathrm{TO{\cdot}CO}\ (0.5 + 0.5F) + \mathrm{ANR} \\
&= 0.078\ \mathrm{TPI} + 2.0\ \mathrm{TO{\cdot}CO}\ (1.0 + F) + \mathrm{ANR}
\end{aligned}
$$

In order to obtain the total annual production cost, the following items must be added to the total net annual operating cost:

1. depreciation
2. average yearly interest on borrowed capital
3. average yearly net return on equity
4. average yearly income tax

The straight-line method was used to determine depreciation, based on the total capital required less the working capital. For stack gas scrubbing units (15 year life), the annual depreciation, ACR, is:

$$
\begin{aligned}
\mathrm{ACR} &= 1/15\ (\mathrm{TCR}-\mathrm{WKC}) \\
&= 0.067\ (\mathrm{TCR}-0.20\ \mathrm{AOC})
\end{aligned}
$$

For substitute natural gas and solvent refined coal plants (20 year life), it is given by:

$$\mathrm{ACR} = 0.050\ (\mathrm{TCR} - 0.20\ \mathrm{AOC})$$

For power plants, both conventional and combined cycle (28 year life), it is:

$$\mathrm{ACR} = 0.036\ (\mathrm{TCR} - 0.20\ \mathrm{AOC})$$

Interest on debt and return on equity are calculated following a procedure recommended in the literature (13) and illustrated in Appendix A. The procedure assumes a fixed debt-to-equity ratio, an interest rate on debt, and the required net (after tax) rate of return on equity. Interest on debt and return on equity are calculated over the plant life, and the yearly average is expressed as a percentage of the total capital required (TCR). Assuming a 75%/25% debt-to-equity ratio, a 9% per year interest rate, and a 15% per year net rate of return on equity, the annual interest and return, AIC, is given by:

$$\text{AIC} = 0.054\ \text{TCR}$$

Federal income tax is the average yearly income tax over the plant life, expressed as a percentage of the total capital required. The calculation of income tax is illustrated in Appendix A. Based on the assumptions listed in the preceding paragraph and an assumed tax rate of 48%, the annual federal income tax, AFT, is given by :

$$\text{AFT} = 0.018\ \text{TCR}$$

The total annual production cost, TAC, is given by:

$$\text{TAC} = \text{AOC} + \text{ACR} + \text{AIC} + \text{AFT}$$

For stack gas scrubbing plants, this can be reduced as follows:

$$\begin{aligned}\text{TAC} &= \text{AOC} + 0.067\ (\text{TCR} - 0.20\ \text{AOC}) + 0.054\ \text{TCR} + 0.018\ \text{TCR}\\ &= \text{AOC} + 0.067\ \text{TCR} - .013\ \text{AOC} + 0.054\ \text{TCR} + 0.018\ \text{TCR}\\ &= 0.139\ \text{TCR} + 0.99\ \text{AOC}\end{aligned}$$

Substituting for TCR and AOC from preceeding equations:

$$\begin{aligned} \mathrm{TAC} &= 0.139\ [1.15\ \mathrm{TPI} + 0.8\ \mathrm{TO}\cdot\mathrm{CO}\ (1.0 + \mathrm{F}) + 0.40\ \mathrm{ANR}] \\ &\quad + 0.99\ [0.078\ \mathrm{TPI} + 2.0\ \mathrm{TO}\cdot\mathrm{CO}\ (1.0 + \mathrm{F}) + \mathrm{ANR}] \\ &= 0.237\ \mathrm{TPI} + 2.1\ \mathrm{TO}\cdot\mathrm{CO}\ (1.0 + \mathrm{F}) + 1.04\ \mathrm{ANR} \end{aligned}$$

Making the appropriate substitutions, the total annual production cost for substitute natural gas and solvent refined coal plants is:

$$\mathrm{TAC} = 0.225\ \mathrm{TPI} + 2.1\ \mathrm{TO}\cdot\mathrm{CO}\ (1.0 + \mathrm{F}) + 1.04\ \mathrm{ANR}$$

For power plants, this equation becomes:

$$\mathrm{TAC} = 0.208\ \mathrm{TPI} + 2.1\ \mathrm{TO}\cdot\mathrm{CO}\ (1.0 + \mathrm{F}) + 1.04\ \mathrm{ANR}$$

The buildup of costs to determine the total annual production cost is illustrated in Figure 4.2.

4.3 Effect of Location on Plant Cost

The cost models have been developed using U.S. Gulf Coast 1973 costs as a basis. In order to predict plant costs for other locations, factors have been developed which relate construction labor costs at various locations to Gulf Coast labor costs. By multiplying the field labor construction portion of plant cost by this location factor, the total plant cost is adjusted to the desired location.

Labor rates for different crafts were obtained from the literature (10) and escalated to the end of 1973. Using an average craft mix obtained from in-house information (12), an average construction labor rate was obtained for each location. Productivity factors for the various locations, also obtained from in-house data, were used to create the rate for equal work output. These rates were then normalized, using Houston (Gulf Coast) as a basis, to yield relative field labor construction costs.

Table 4.1 lists the relative labor costs determined for twenty cities. They range from 1.0 for Houston to 2.08 for New York. Costs are generally highest in the Northeastern quarter of the country and lowest in the South. These factors are shown on a map of the U.S. in Figure 4.3.

Table 4.2 lists average location factors for each state. Allowance has been made in the factor for the importation of temporary labor to the more remote states. The factors are shown on a map of the U.S. in Figure 4.4.

Figure 4.5 gives the relationship between major equipment cost, E, total plant investment, TPI, and location factor, F, when the contingency, CONTIN, is zero.

4.4 Nomenclature

E	Major equipment costs	M$
M	Other material costs	M$
L	Direct field labor costs (Gulf Coast)	M$
BARC	Bare cost	M$
F	Location factor	
CONTIN	Contingency	
TPI	Total plant investment	M$
STC	Start-up costs	M$
WKC	Working capital	M$
IDC	Interest during construction	M$
TCR	Total capital required	M$
ANR	Annual cost of raw materials, utilities, and waste disposal, less by-product credits	M$/year
AOL	Annual cost of operating labor and supervision	M$/year
AML	Annual cost of maintenance labor and supervision	M$/year
APS	Annual cost of plant supplies and replacements	M$/year

AOH	Annual cost of administration and overheads	M$/year
ATI	Annual cost of local taxes and insurance	M$/year
AOC	Total net annual operating cost	M$/year
TO	Total number of shift operators	
CO	Hourly rate for shift operators (Gulf Coast)	$/hour
ACR	Annual depreciation	M$/year
AIC	Annual interest on debt and return on capital	M$/year
AFT	Annual federal income taxes	M$/year
TAC	Total annual production cost	M$/year
COHP	Contractor overhead & profits	M$/year
TAXI	Taxes and insurance	M$/year
FLIC	Field Labor Indirect Cost	M$/year
ENGR	Engineering Fees	M$/year

TABLE 4.1

LOCATION FACTORS FOR MAJOR U.S. CITIES

Location	Location Factor F
Atlanta	1.10
Baltimore	1.41
Birmingham	1.16
Boston	1.23
Chicago	1.52
Cincinnati	1.53
Cleveland	1.86
Dallas	1.07
Denver	1.03
Detroit	1.73
Kansas City	1.37
Los Angeles	1.44
Minneapolis	1.54
New Orleans	1.16
New York	2.08
Philadelphia	1.82
Pittsburgh	1.52
St. Louis	2.01
San Francisco	1.45
Seattle	1.21
Houston	1.00

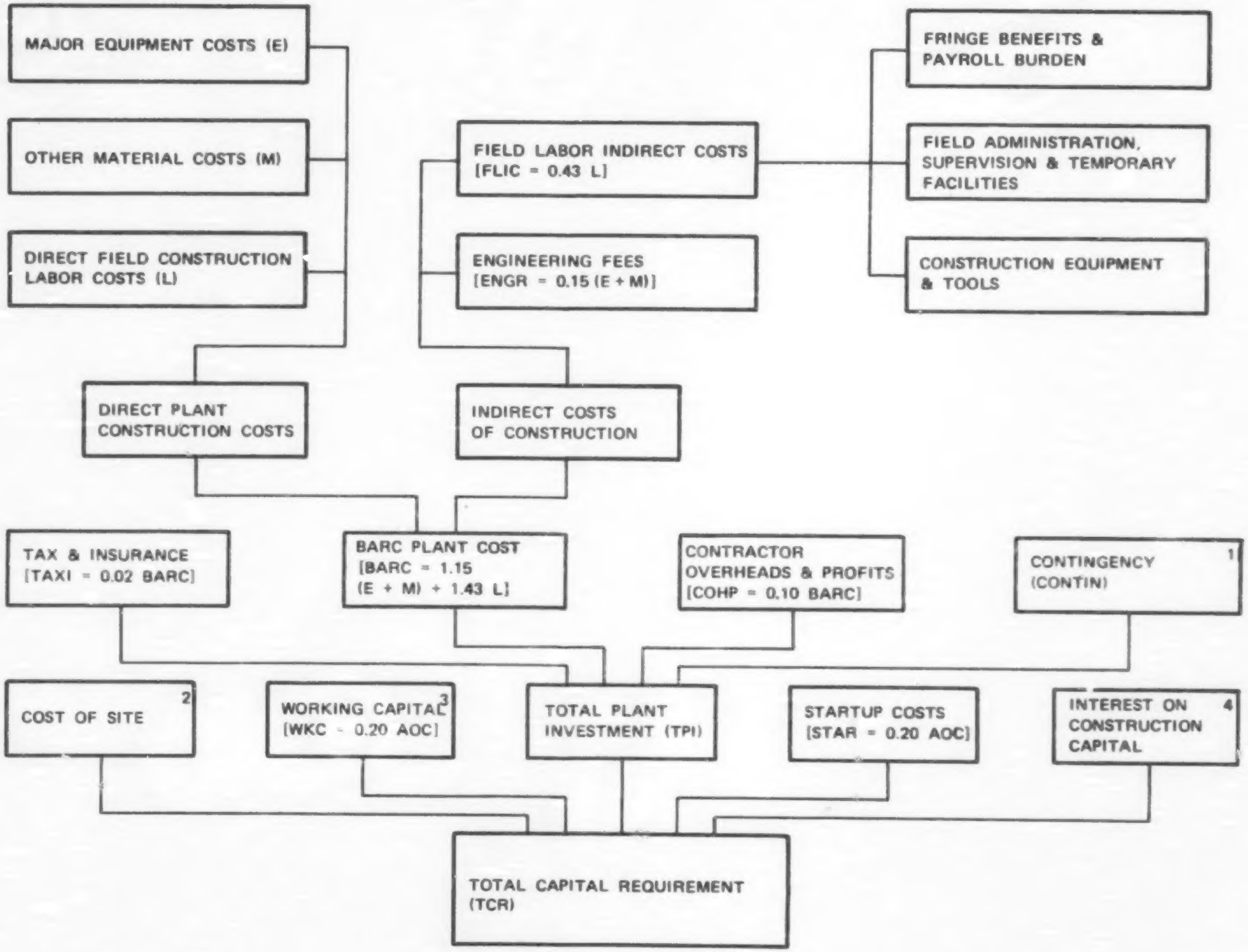
TABLE 4.2

AVERAGE LOCATION FACTORS FOR EACH STATE

State	Location Factor
Alabama	1.2
Alaska	2.1
Arizona	1.3
Arkansas	1.2
California	1.5
Colorado	1.2
Connecticut	1.7
Delaware	1.4
D.C.	1.4
Florida	1.2
Georgia	1.1
Hawaii	2.0
Idaho	1.3
Illinois	1.7
Indiana	1.6
Iowa	1.5
Kansas	1.4
Kentucky	1.5
Louisiana	1.1
Maine	1.2
Maryland	1.4
Massachusetts	1.3
Michigan	1.7
Minnesota	1.5
Mississippi	1.1
Missouri	1.6
Montana	1.3
Nebraska	1.4
Nevada	1.4
New Hampshire	1.2
New Jersey	2.1
New Mexico	1.3
New York	2.1
N. Carolina	1.2
North Dakota	1.3
Ohio	1.6
Oklahoma	1.4
Oregon	1.2
Pennsylvania	1.6
Rhode Island	1.3
S. Carolina	1.1
South Dakota	1.3
Tennessee	1.2
Texas	1.1
Utah	1.2
Vermont	1.2
Virginia	1.4
Washington	1.2
W. Virginia	1.5
Wisconsin	1.5
Wyoming	1.3

FIGURE 4.1

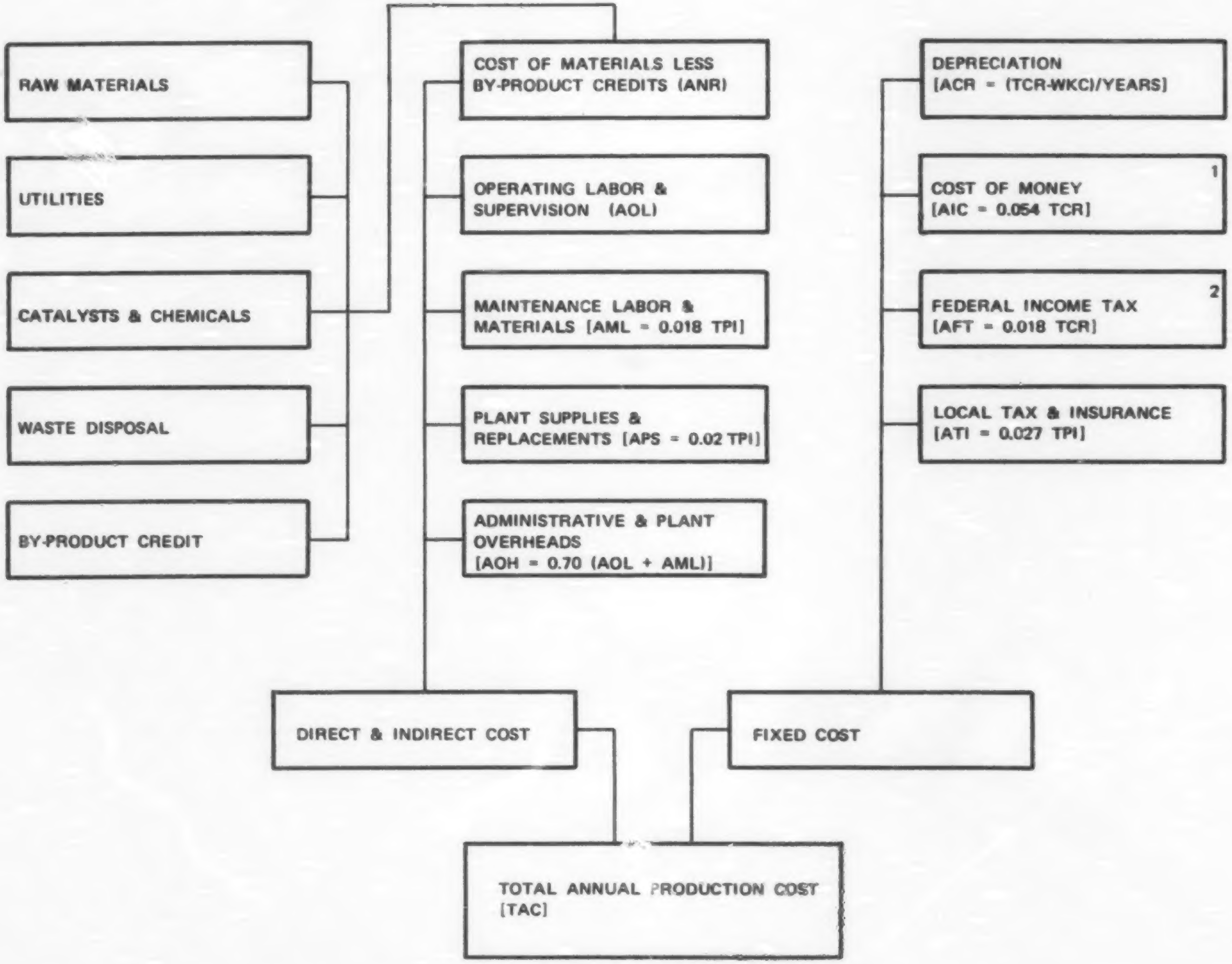
RELATIONSHIP BETWEEN CAPITAL COST FACTORS IN THE GENERAL COST MODEL


MAJOR EQUIPMENT COSTS (E)
OTHER MATERIAL COSTS (M)
DIRECT FIELD CONSTRUCTION LABOR COSTS (L)
FIELD LABOR INDIRECT COSTS [FLIC = 0.43 L]
ENGINEERING FEES [ENGR = 0.15 (E + M)]
FRINGE BENEFITS & PAYROLL BURDEN
FIELD ADMINISTRATION, SUPERVISION & TEMPORARY FACILITIES
CONSTRUCTION EQUIPMENT & TOOLS
DIRECT PLANT CONSTRUCTION COSTS
INDIRECT COSTS OF CONSTRUCTION
TAX & INSURANCE [TAXI = 0.02 BARC]
BARC PLANT COST [BARC = 1.15 (E + M) + 1.43 L]
CONTRACTOR OVERHEADS & PROFITS [COHP = 0.10 BARC]
CONTINGENCY (CONTIN) 1
COST OF SITE 2
WORKING CAPITAL 3 [WKC - 0.20 AOC]
TOTAL PLANT INVESTMENT (TPI)
STARTUP COSTS [STAR = 0.20 AOC]
INTEREST ON CONSTRUCTION CAPITAL 4
TOTAL CAPITAL REQUIREMENT (TCR)

1. SEE DEFINITION ON PAGE 58.
2. COST WOULD NORMALLY BE INCLUDED ONLY IF PURCHASE IS REQUIRED. COST IS USUALLY SMALL AND HAS NOT BEEN INCLUDED IN MODEL.
3. SEE NOTE 3 OF FIGURE 4.2.
4. SEE FIGURE 4.2.

FIGURE 4.2

RELATIONSHIP BETWEEN PRODUCTION COST FACTORS IN THE GENERAL COST MODEL


RAW MATERIALS
UTILITIES
CATALYSTS & CHEMICALS
WASTE DISPOSAL
BY-PRODUCT CREDIT
COST OF MATERIALS LESS BY-PRODUCT CREDITS (ANR)
OPERATING LABOR & SUPERVISION (AOL)
MAINTENANCE LABOR & MATERIALS [AML = 0.018 TPI]
PLANT SUPPLIES & REPLACEMENTS [APS = 0.02 TPI]
ADMINISTRATIVE & PLANT OVERHEADS [AOH = 0.70 (AOL + AML)]
DEPRECIATION [ACR = (TCR-WKC)/YEARS]
COST OF MONEY [AIC = 0.054 TCR] 1
FEDERAL INCOME TAX [AFT = 0.018 TCR] 2
LOCAL TAX & INSURANCE [ATI = 0.027 TPI]
DIRECT & INDIRECT COST
FIXED COST
TOTAL ANNUAL PRODUCTION COST [TAC]

1. AVERAGE OVER THE PLANT LIFE, ASSUMING 75% DEBT AT 9% INTEREST RATE PER YEAR, AND 25% EQUITY GIVING A NET RETURN OF 15%.
2. AVERAGE OVER THE PLANT LIFE, ASSUMING 48% FEDERAL INCOME TAX RATE.
3. ANNUAL OPERATING COST IS: AOC = ANR + AOL + AML + APS + AOH + ATI.

FIGURE 4.3

LOCATION FACTORS FOR SELECTED CITIES

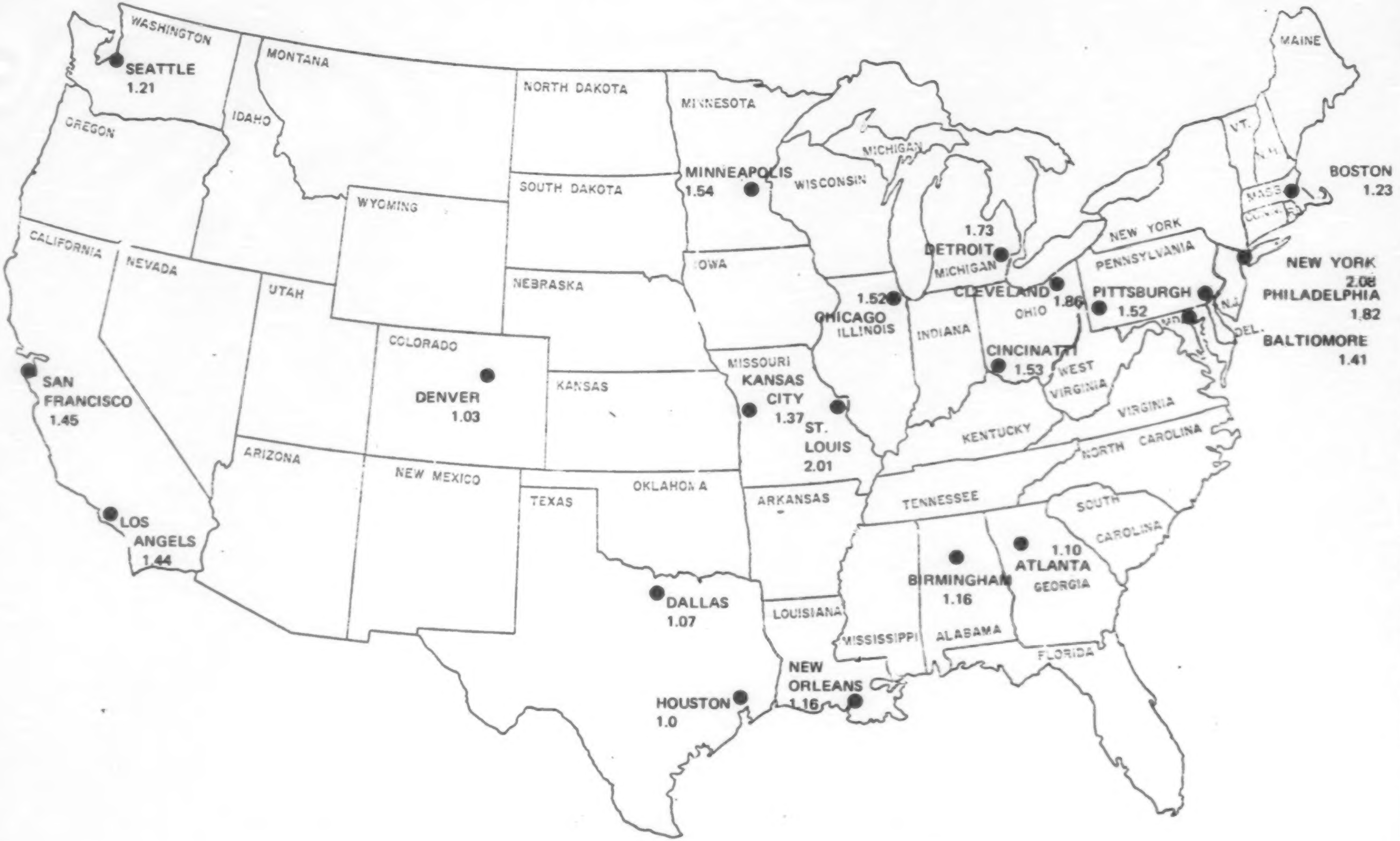

SEATTLE 1.21
SAN FRANCISCO 1.45
LOS ANGELS 1.44
DENVER 1.03
DALLAS 1.07
HOUSTON 1.0
MINNEAPOLIS 1.54
KANSAS CITY 1.37
ST. LOUIS 2.01
NEW ORLEANS 1.16
CHICAGO 1.52
DETROIT 1.73
CLEVELAND 1.86
CINCINATTI 1.53
BIRMINGHAM 1.16
ATLANTA 1.10
PITTSBURGH 1.52
BOSTON 1.23
NEW YORK 2.08
PHILADELPHIA 1.82
BALTIMORE 1.41
WASHINGTON
OREGON
CALIFORNIA
NEVADA
IDAHO
MONTANA
UTAH
ARIZONA
WYOMING
COLORADO
NEW MEXICO
NORTH DAKOTA
SOUTH DAKOTA
NEBRASKA
KANSAS
OKLAHOMA
TEXAS
MINNESOTA
IOWA
MISSOURI
ARKANSAS
LOUISIANA
WISCONSIN
ILLINOIS
MICHIGAN
INDIANA
OHIO
KENTUCKY
TENNESSEE
MISSISSIPPI
ALABAMA
GEORGIA
FLORIDA
SOUTH CAROLINA
NORTH CAROLINA
VIRGINIA
WEST VIRGINIA
PENNSYLVANIA
NEW YORK
N.J.
DEL.
MD.
MASS
CONN.
VT.
N.H.
MAINE

FIGURE 4.4

AVERAGE LOCATION FACTORS BY STATE

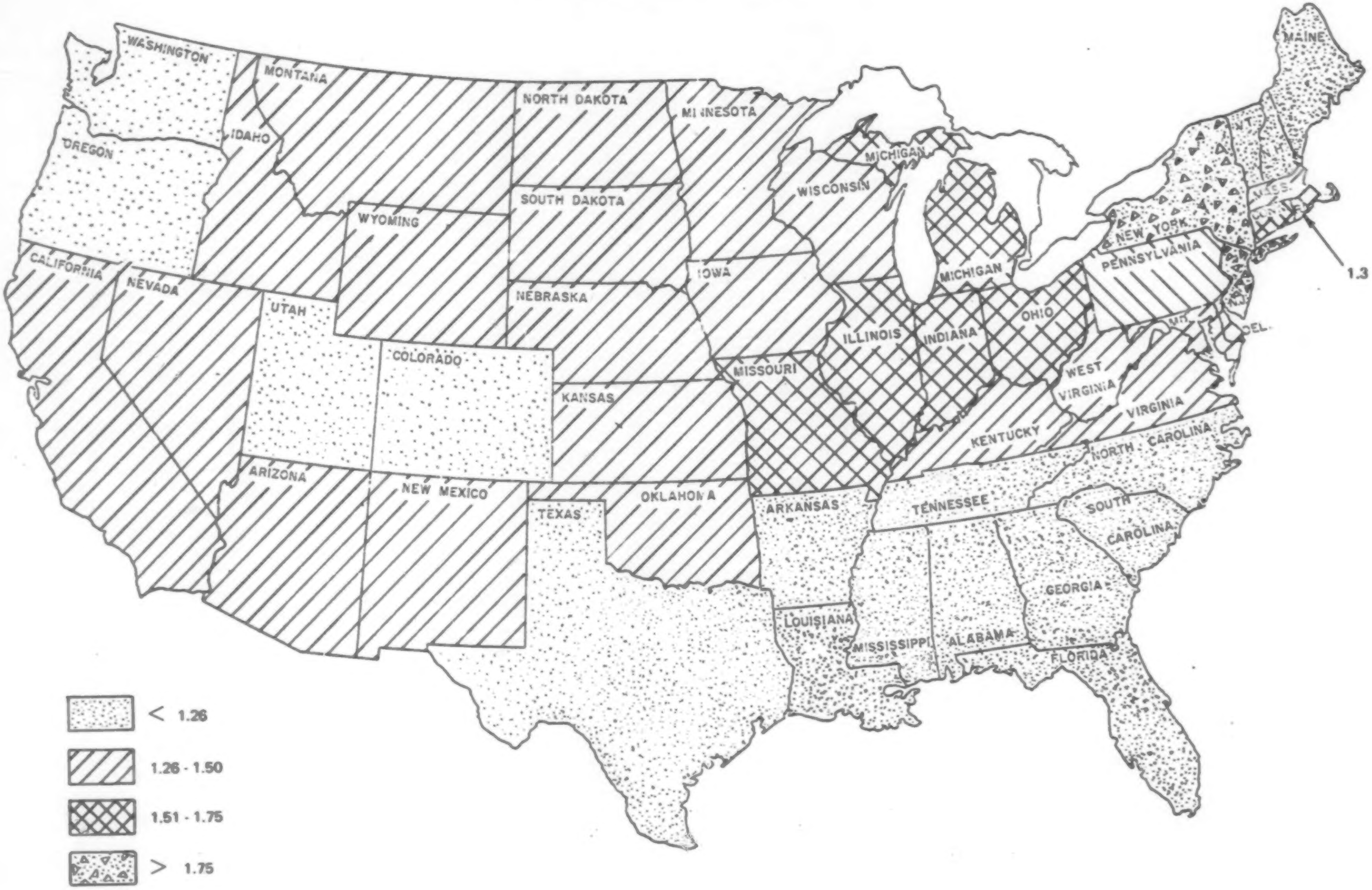
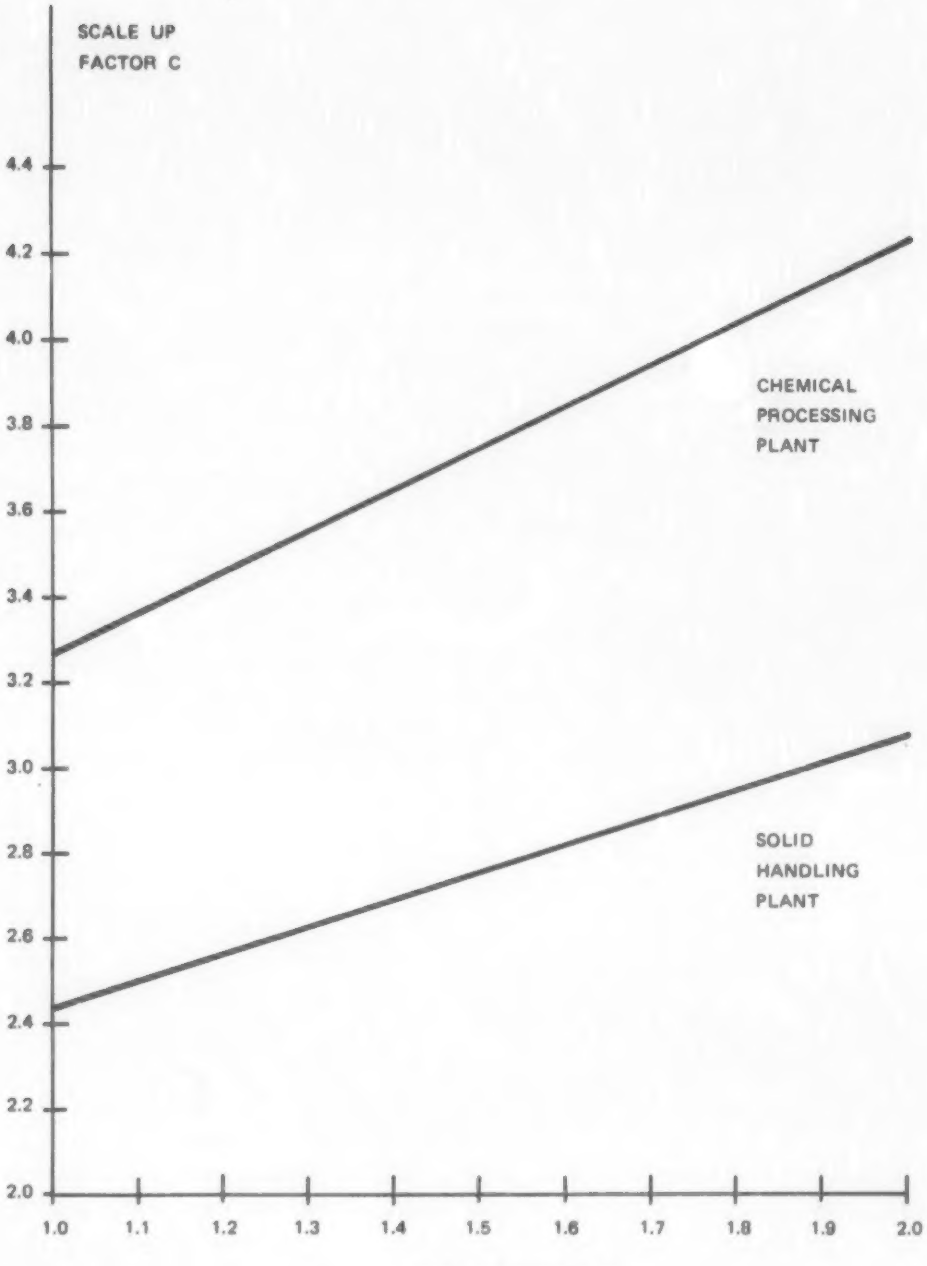

WASHINGTON
OREGON
CALIFORNIA
IDAHO
NEVADA
MONTANA
WYOMING
UTAH
ARIZONA
COLORADO
NEW MEXICO
NORTH DAKOTA
SOUTH DAKOTA
NEBRASKA
KANSAS
OKLAHOMA
TEXAS
MINNESOTA
IOWA
MISSOURI
ARKANSAS
LOUISIANA
WISCONSIN
MICHIGAN
ILLINOIS
INDIANA
MICHIGAN
OHIO
KENTUCKY
TENNESSEE
MISSISSIPPI
ALABAMA
GEORGIA
FLORIDA
SOUTH CAROLINA
NORTH CAROLINA
VIRGINIA
WEST VIRGINIA
PENNSYLVANIA
NEW YORK
MD.
DEL.
N.J.
VT.
N.H.
MASS.
MAINE
1.3
< 1.26
1.26 - 1.50
1.51 - 1.75
> 1.75

FIGURE 4.5

EFFECT OF LOCATION FACTOR ON TOTAL PLANT INVESTMENT (CONTINGENCY = 0)

$TPI = C \cdot E$



5. THE WET LIMESTONE PROCESS

5.1 Process Appraisal

The wet limestone process has been re-examined in the light of the experiences of the Will County Plant, the Shawnee Plant and the estimate study carried out for EPA by Catalytic Inc. (16) Several process alternatives were considered for various sections of the plant and the conclusion was that the process flowsheet presented by Catalytic Inc. was the best available solution for the wet limestone process.

Heat recovery from the flue gas can be dismissed as a practical possibility. The flue gas would be cooled below its acid dew point and consequently expensive alloy exchangers would be required. In order to achieve the required exchanger area, finned tubes would be required. These would soon be blocked by wet fly ash. The most sensible method of cooling is obviously by direct scrubbing in a venturi. The slurry temperature is allowed to rise to 130°F and all the heat is removed in saturating the flue gas at this temperature.

A turbulent contact absorber has been selected to handle the limestone slurry which both silts up and scales equipment. This is also the design used by Catalytic Inc. In view of the operating difficulties experienced by Will County with blockage of the Chevron demister, it appears sensible to make this separate from the TCA and put considerable thought into its design.

The clean flue gas must be reheated from 130°F to 200°F to restore its buoyancy and reduce its relative humidity, so there is no alternative but to provide this heat by burning extra fuel. Although this could amount to 1 or 2% of the boiler fuel, it is still the most sensible and economic design. The in-line burner appears a better idea than indirect heating with steam or air, since the direct heat exchange is more efficient and the additional equipment required is less. A small quantity of low sulfur ash free fuel is required for this purpose.

Because of the limited nature of the results at this point in time it is desirable to opt for a safe design. This appears to be a slurry flowrate to the venturi of 11 gpm/MACFM inlet gas and a slurry flowrate to the TCA of 55 gpm for the same quantity of gas. The pressure drop in the venturi should be about 10 inch w.g. and the gas velocity in the TCA 10 ft/sec. With a venturi and a 2 stage TCA, the absorption efficiency at 2500 ppm SO_2 inlet is about 87%, and with a venturi and 3 stage TCA the efficiency is about 90%. Although 90% absorption would only be required when burning 6 or 7% sulfur coals there is little point in designing for anything less than 90%; the slurry flowrate could not be significantly reduced and the cost of an additional TCA stage is relatively small. So a venturi and a 3 stage TCA with the above flowrate will be a constant unit in the wet limestone model. The process then has the flexibility in the scrubbing section to clean up flue gas to within the federal limit of 1.2 lb SO_2/MMBtu even when burning 6-7% sulfur coal. A change to a coal containing more sulfur could be easily accomodated by installing additional limestone slurry preparation and waste removal units. Short term changes could even be handled by using the spare limestone grinding and slurrying equipment.

Two changes have been made to the Catalytic design:

1. The slurry flowrate to the TCA was doubled and the number of stages increased from 2 to 3.

2. The differential produced by the induction fan was increased from 18 to 30 inch w.g. which is made up of the following:

	inches w.g.
Venturi	10
3 stage TCA	8
Entrainment separator	2
New ductwork	7
Plugging allowance	3
	30

The maximum sized venturi and TCA unit will handle a gas flow to the venturi of 550,000 ACFM. This flowrate corresponds to approximately one sixth of the total gas flow rate at 300°F from a 1000 megawatt coal fired power station with a heat rate of 9650 Btu/kwh, assuming a 10% increase due to leakage in the air preheater.

The process flowsheet is shown in Figure 5.1. The process has been divided into 3 basic sections:

1. The scrubbing system, including flue gas reheat and discharge
2. The limestone handling and slurry preparation section
3. The waste disposal system including the settling pond.

The first section consists mainly of parallel trains of equipment. The reference cost for each item of equipment in a train will be for the size which handles 550,000 ACFM flue gas to the venturi. The reference cost for equipment in the scrubbing section which is not required in parallel trains will be the size handling 3,300,000 ACFM flue gas to the venturis. This corresponds approximately to the size of a 1000 megawatt facility.

The second and third sections will be in one train whatever the size of the wet limestone facility. Costs in these sections will be referenced to a sulfur flow rate in the fuel burned of 28,000 lb/hr, or approximately a 1000 megawatt power station burning a 4% sulfur coal.

5.2 Evaluation of Catalytic Inc. Estimate

A general examination of the Catalytic Inc. estimate by the MWK Estimating Department showed it to be sound. A breakdown of the total plant cost agrees closely with the equivalent MWK figures. A close examination of all equipment material and subcontract costs and comparison with quotations made to MWK shows them to be reasonable, with the exception of the induction fans and motors.

The cost of a 1500 BHP, 380,000 ACFM fan and motor for the TVA Gallatin 1050 megawatt plant built in 1955 was $81,000 and so the Catalytic figure of $90,000 (end of 73) for a 3000 BHP 360,000 ACFM I.D. fan and motor appears to be much too low. Quotations made recently to MWK in fact show this to be so, the cost of a 4,500 BHP 475,000 ACFM I.D. fan and motor was $210,000 (end of 73). For the maximum sized unit 550,000 ACFM (almost the same ACFM as to the venturi inlet), the BHP will be about 4,000. So a safe assumption for the cost of the incremental I.D. fan and motor appears to be $200,000 (end of 73).

The basis for all other equipment costs is the Catalytic Inc. figure listed in Appendix B.

5.3 Variation of Equipment Costs with Plant Size

An article by K.M. Guthrie in the March 1969 issue of Chemical Engineering, "Data and Techniques for Preliminary Capital Cost Estimating" (9), has been used to establish how equipment costs vary with size.

For most types of equipment, the cost does not vary with size in the same way over the whole size range. For example the cost of centrifugal pumps and motors varies with the 0.4 power of the BHP over the lower size range and the power increases to 0.6 for larger machines. This can make cost prediction from one quotation using one exponent very inaccurate especially where large variations in size are involved.

Referring to Figure 5.2, the curve shown represents the variation of equipment cost with size. This curve is not an easy thing to establish and various sources of information have different ideas on the relationship between equipment cost and size. However assuming it to be available and accurate, it is still difficult to use it safely in simple form, when the quotations could be for equipment on any part of the curve.

If the equipment cost which is to be used as a basis for the cost equation is in the middle of the curve, then using the average exponent results in seriously underestimating the cost of larger and smaller equipment. A safer method is illustrated by considering the example of the centrifugal pump and motor in Figure 5.2.

The known equipment cost (\$5,000 for 100 BHP) is scaled up to the cost for the maximum required size (200 BHP) using the exponent for the higher end of the curve 0.6. This maximum cost (\$7,600) is used in the equipment cost equation which is then based on the average exponent of 0.5. Thus:

$$\text{Pump/Motor Cost} = \$7{,}600\left(\frac{\text{BHP}}{200}\right)^{0.5}$$

In this way the estimated value is always greater than or equal to the cost curve, which is better in view of the accuracy of the exponents. However, this method should be used with care, keeping the overall equipment size range to about 2.0:1.

If the known equipment cost is for a size much smaller than the required maximum size, then it would be preferable to obtain another cost estimate closer to the maximum size. In lieu of this, however, the average exponent could be used for both scaling up and scaling down.

The equipment required in this process is listed below with the exponent relating cost to size.

	Cost Proportional to
Distillation and absorber tower shells (this applies to the venturi, TCA, and ductwork)	$(\mathrm{ACFM})^{0.4-0.6}$
Distillation column trays (this applies to the entrainment separator)	$(\mathrm{ACFM})^{0.8-1.0}$
Centrifugal pumps and agitators	$(\mathrm{BHP})^{0.4-0.6}$
Horizontal pressure vessels - constant pressure	$(\mathrm{volume})^{0.4-0.6}$
Storage tanks - up to 200,000 gal	$(\mathrm{volume})^{0.4-0.6}$
Storage silos	$(\mathrm{volume})^{0.8-1.0}$
Conveyors and feeders (length and height fixed, quantity handled varies)	$(\mathrm{quantity})^{0.8-1.0}$
Tube mill wet grinders	$(\mathrm{quantity})^{0.8-1.0}$
Separating ponds	$(\mathrm{quantity})^{0.8-1.0}$

The total flue gas flow rate to the venturis in the Catalytic Inc. design is 1,520,000 ACFM and there are 4 scrubbing trains which handle 380,000 ACFM each. The total sulfur flow into the control process is 13,000 lb/hr. Therefore the factors used to scale up the Catalytic's costs to the maximum size unit are:

$\frac{550}{380} = 1.45$ $(1.45)^{0.6} = 1.25$

$\frac{3{,}300}{1{,}520} = 2.17$ $(2.17)^{0.6} = 1.59$

$\frac{28}{13} = 2.16$ $(2.16)^{0.6} = 1.58$

5.4 Cost Model

5.4.1 Equipment Costs

Using Catalytic's estimate and the scale-up factors given in the preceding section, equipment costs have been calculated for the maximum or reference size units and are shown below.

1. The Scrubbing System

	Cost of Maximum Size Train M$ (end of 73)	Cost Relationship with GT
A. Venturi, 3 stage TCA (extra stage, 15%) and sumps 1.25 [273 + (908 x 1.15) + 274]/4	= 499	$GT^{0.5}$
B. Entrainment Separator 1.45 (574)/4	= 208	$GT^{0.9}$
C. Venturi recirculation tank, agitator and pumps 1.25 [92 + 28.5 + 51.5]/4	= 54	$GT^{0.5}$
D. TCA recirculation tank, agitator and pumps 1.25 [131 + 33.5 + (98.3 x 1.41)]/4	= 95	$GT^{0.5}$
E. Ductwork (including dampers) and reheater 1.25 (1,085 + 169)/4	= 393	$GT^{0.5}$
F. I.D. fans and motors (incremental for control facility)	= 200	$GT^{0.9}$

The cost of equipment in each scrubbing & reheat train

$= 1041\ (GT/550)^{0.5} + 408\ (GT/550)^{0.9}$ in M$ where GT is the gas flow rate to each train in MACFM.

In addition to the equipment in the parallel trains there is the emergency ammonia injection system (G), the entrainment separator recirculation system (H), the reheat fuel storage and delivery system (J).

The total cost of these three units for the reference size plant is

$1.59\ (10.8 + 64.2 + 75) = 238$ M$

The cost of these three units for a plant handling a total gas flowrate of GP MACFM

$= 238 \left(\frac{GP}{3,300}\right)^{0.5}$ M$

2. The Limestone Handling and Slurry Preparation System

	Cost of Reference Size Unit M$ (end of 73)	Cost Relationship with SF
K. Limestone silo conveyor and stockpile feeder 2.16 (61.7)	= 133	$SF^{0.9}$
L. Limestone silo and feeders 2.16 (82 + 23.1)	= 227	$SF^{0.9}$
M. 3 tube mill wet grinders and air compressor 2.16 (595 + 13.5)	=1320	$SF^{0.9}$

	Cost of Reference Size Unit M$ (end of 73)	Cost Relationship with SF
P. Slurry hold up tanks, agitators & feed pumps 1.58 (1.6 + 50 + 29.1 + 2.4 + 3.2)	= 136	$SF^{0.5}$

The cost of all the equipment in the limestone handling system

$$= 1680\ (SF/28)^{0.9} + 136\ (SF/28)^{0.5} \qquad M\$$$

where SF is the total sulfur flow into the control unit in M lb/hr.

3. The Waste Disposal System Including the Settling Pond

	Cost of Reference Size Unit M$ (end of 73)	Cost Relationship with SF
Q. Surge tanks and pumps serving pond 1.58 (6.7 + 34.5)	= 65	$SF^{0.5}$
R. Separating Pond (80% load factor) 0.7 x 2.16 (4,000) NB The Catalytic design included approx. 30% space for fly ash, which is not an SO_x control cost. The cost of the separating pond represents the cost in Cincinnati.	= 6,000	$SF^{0.9}$

The cost of the surge tanks and pumps serving the pond

$$= 65\ (SF/28)^{0.5} \qquad M\$$$

The cost of the separating pond

$$= 6{,}000\left[\frac{SF.LF}{28\times 0.8}\right]^{0.9} = 6{,}000\left[\frac{SF.LF}{22}\right]^{0.9} \qquad M\$$$

Where LF is the load factor.

The total equipment cost (material and subcontract) for chemical processing plant (EC) in the Wet Limestone process

$$= \sum_{n=1}^{NA} RB\,[1041\,(GT/550)^{0.5} + 408\,(GT/550)^{0.9}]_n$$

$$+ 238\,RP\,(GP/3{,}300)^{0.5} + 201\,(SF/28)^{0.5} \qquad M\$$$

where NA is the number of scrubbing trains and RB **and** RP are retrofit difficulty factors as explained below.

The total equipment cost for solid handling plant (ES)

$$= 1680\,(SF/28)^{0.9} \qquad M\$$$

In addition to this, the material and construction costs of the reference size separating pond, adjusted to 100% load factor and Gulf Coast location is $5,000 M. Thus:

$$P = 5{,}000\left[\frac{SF\cdot LF}{28}\right]^{0.9} \qquad M\$$$

RB is the retrofit difficulty factor of the individual boiler. The increased difficulty is not so much reflected in the actual major equipment costs as in the increase in other material and labor costs associated with them. However this is a convenient place to introduce the factor.

RP is the retrofit difficulty factor of the rest of the scrubbing section which is not in parallel trains. This has been assumed equal to the highest RB in the plant.

An examination of the MWK reports for EPA, "Applicability of SO_2 Control Processes to Power Plants" and "Evaluation of the Controllability of Power Plants Having a Significant Impact on Air Quality Standards" (17, 18), produces a simplified table of boiler retrofit factors:

TABLE 5.1 Boiler Retrofit Factors

Boiler Size (Megawatts)	Boiler Age (Years)	RB
< 50	>10	2.0
	<10	1.8
50-100	>10	1.8
	<10	1.6
101-200	All	1.6
201-500	All	1.5
>500	All	1.4
All new boilers	-	1.0

5.4.2 Other Material Costs and Labor Costs

The Guthrie paper (9) indicates that different relationships exist between major equipment costs, other material costs and labor costs for chemical process plant and for solid handling plant. This was found to be true for the Catalytic Inc. estimate although the relationships did not agree with the Guthrie paper. This is not really surprising as it depends on how the job is contracted out and estimated. The obvious solution is to use the relationships generated from the Catalytic figures since they will be used with Catalytic's major equipment costs. These costs are listed in Appendix C.

Major Equipment Costs, E:

$$E = EC + ES$$

Field Labor Costs, L (U.S. Gulf Coast):

$$LC = 0.39\ EC$$
$$LS = 0.18\ ES$$
$$L = LC + LS$$

Other Material Costs, M:

$$MC = 0.82\ EC$$
$$MS = 0.09\ ES$$
$$M = MC + MS$$

The letter C after the letters E, L and M denotes chemical process type plant. The letter S denotes solid handling plant.

5.4.3 Raw Material and Utilities Costs

1. Limestone

The quantity of limestone used by the process during the year is directly proportional to the sulfur flow into the control unit, SF, and the boiler load factor, LF. The Catalytic plant uses 32 tons/hr of limestone for a sulfur flowrate into the control unit of 13,000 lb/hr. The reference flowrate is 28,000 lb/hr.

The limestone used for the reference flow at 100% load factor

$$= 32 \times \frac{28}{13} \times 8760 = 600\ \text{M tons/year}$$

the cost of limestone, $A1 = 600\ CL \cdot LF\ (SF/28)$ M$

where CL is the purchase price of limestone, $/ton.

2. Ammonia

Ammonia is used intermittently and the yearly consumption is estimated to be 200 tons. The number of upsets requiring ammonia injection probably will not reduce with reduction of the load factor, since startups and shutdowns represent unsteady conditions. Since the cost is small, for simplification the use of ammonia will be assumed directly proportional to the sulfur flow into the control unit.

Ammonia used for reference flow

$$= 200 \times \frac{28}{13} = 0.43 \text{ M tons/year}$$

The cost of ammonia, AA = 0.43 CA (SF/28) M$/year

where CA is the purchase price of ammonia, $/ton.

3. Process Water

The consumption of process water is 400 gpm, which is lost almost equally between the settling pond and the exhausting flue gas.

The scale-up factors for the reference flows are:

$$\frac{3300}{1520} = 2.17 \text{ and } \frac{28}{13} = 2.16$$

The water consumption per year at the reference flowrate and 100% load factor

$$= 2.17 \times 400 \times 60 \times 8760$$

$$= 460{,}000 \text{ M Gal/year}$$

The cost of process water,

$$AW = 230\ CW \cdot LF\ [(GP/3{,}300) + (SF/28)] \qquad M\$/year$$

where CW is the purchase price of water, $/M Gal.

4. Fuel Oil

The consumption of fuel oil in the catalytic design is 95 MMBtu/hr.

The consumption of fuel oil at the reference flowrate and 100% load factor

$$= 2.17 \times 95 \times 8760 \qquad MMBtu/year$$
$$= 1{,}800{,}000 \qquad MMBtu/year$$

The cost of fuel oil,

$$AF = 1{,}800\ CF \cdot LF\ (GP/3{,}300) \qquad M\$/year$$

where CF is the purchase price of fuel oil, $/MMBtu.

5. Electricity

The electricity used in the scrubbing section has been increased by 1100 kw to cover doubling the slurry flow to the TCA. The rating of the flue gas fans has been increased by 4300 kw to cover the increased pressure differential. The total electricity consumed is now 13,050 kw. Of this 11,210 kw are proportional to GP and 1,840 kw are proportional to the sulfur flow.

The electricity consumption per year at the reference flowrates and 100% load factor

$$= 2.17 \times 11.21 \times 8760 \text{ (proportional to GP)}$$
$$= 213{,}000 \text{ M. kwh}$$

and

$$2.17 \times 1.840 \times 8760 \text{ (proportional to SF)}$$
$$= 35{,}000 \text{ M. kwh}$$

The cost of electricity,

$$AE = CE \cdot LF\,[213\,(GP/3{,}300) + 35\,(SF/28)] \qquad \text{M\$/year}$$

where CE is the purchase price of electricity, mils/kwh.

The total incremental energy consumption of the Wet Limestone scrubbing unit amounts to about 5% of the HHV of the coal feed to the power plant.

The total annual cost of raw materials and utilities, ANR, is given by:

$$ANR = AL + AA + AW + AF + AE$$

5.4.4 Total Plant Investment and Total Capital Required

The main costs of the separating pond (P) are the construction labor costs and land cost and have been assumed to be dependent on the location at which the Wet Limestone unit is to be built. The bare cost of the unit can be derived from the General Cost Model.

$$BARC = 1.15\,(E + M) + (P + 1.43\,L)\,F$$

The Total Plant Investment is given by:

$$TPI = 1.12\,(1.0 + CONTIN)\,BARC$$

The contingency CONTIN, represents the degree of uncertainty in the process design and the cost estimate.

The Total Capital Required is given by the appropriate equation in the General Cost Model.

$$TCR = 1.15\ TPI + 0.8\ TO \cdot CO\ (1 + F) + 0.4\ ANR$$

5.4.5 Operating Costs

The total net annual operating cost, AOC, is the total cost of operating the plant less the credits from the sale of by-product. It does not include return of capital, payment of interest or income tax on equity return. The total net annual operating cost for the Wet Limestone process is given by:

$$AOC = 0.078\ TPI + 2 TO \cdot CO\ (1 + F) + ANR$$

The total number of shift operators, TO, for the Wet Limestone process is 8 (2 men per shift) for plant capacities of 200 megawatts or above. For plants below 200 megawatts, the cost for operating labor is assumed to decrease linearly with size. The hourly wage of the operators, CO, is expressed in $/hr.

The Total Annual Production Cost, TAC, including the return of capital, payment on interest and income tax on equity return is given by:

$$TAC = 0.237\ TPI + 2.1\ TO \cdot CO\ (1 + F) + 1.04\ ANR$$

5.5 Effect of Various Parameters on Costs

In Figures 5.3-5.7 typical costs which were calculated from the model have been plotted to illustrate the effects of different variables on plant costs. Unit values for raw materials and utilities, which were used in determining operating costs, are as shown in Table 5.2

These plots are not for actual, existing plants, but have been included merely to illustrate typical cost variations predicted by the model. Although the figures are self-explanatory, some of the more significant conclusions should be noted.

Figure 5.3 shows the large effect of plant capacity (i.e., gas flow) on capital required. Small plants are far more expensive to control then large ones. While a new 1000 MW plant (4% S, 80% load factor) could be controlled for about $46/KW, for a 10 MW plant it would cost almost three times as much. The sulfur content of the coal has a noticeable but minor effect on cost, particularly at small plant capacities.

Figure 5.4 illustrates the pronounced effect of load factor on operating cost. In fact, decreasing the load factor from 80% to 40% is more significant than quadrupling the sulfur content of the coal. Plant capacity has an effect on operating cost similar to that on capital required.

Figures 5.5-5.6 show the influence of the retrofit factor on costs. As it becomes more difficult to install a wet limestone unit at an existing plant, capital required increases substantially. For a 10 MW plant, it could be more than $250/KW. Even for a large 1000 MW plant, capital required could be as much as 70% more than for a new plant. The increase in operating cost is due to the fixed charges on the additional capital.

In figures 5.7, the effect of location factor on capital required is shown. Basically, this shows the influence of higher labor rates on the construction cost of the plant. Relative to a Gulf Coast location, costs could be as much as 25-35% higher at other locations.

5.6 Nomenclature

GP	Total gas flow into all venturis	MACFM
GT	Total flow of gas into each venturi (Maximum value of GT = 550)	MACFM
NA	Number of venturi/TCA trains (GT = GP/NA for a new plant)	
SF	Maximum flow of sulfur into the control unit	M lb/hr
LF	Load factor of the power station	
RB	The retrofit difficulty factor of a boiler	
RP	The retrofit difficulty factor of all scrubbing equipment which is not in parallel trains. Assumed to be equal to the highest RB	
CL	The purchase price of limestone	$/ton
CA	The purchase price of ammonia	$/ton
CW	The purchase price of process water	$/M Gal
CF	The purchase price of fuel oil	$/MM Btu
CE	The purchase price of electricity	Mils/kwh

CO	The direct cost of operating labor	\$/hour
E	Major equipment cost (Material and subcontract)	M\$
M	Other material costs (Piping, instruments, electrical civil etc.)	M\$
L	Direct field labor costs	M\$
C, S	Letters follows E, M and L C refers to chemical process type equipment S refers to solid handling equipment	
P	The total cost of the settling pond (Material and total labor)	M\$
AL	Total annual cost of limestone	M\$/year
AA	Total annual cost of ammonia	M\$/year
AW	Total annual cost of process water	M\$/year
AF	Total annual cost of fuel oil	M\$/year
AE	Total annual cost of electricity	M\$/year
TPI	The total plant investment	M\$
TAC	Total annual production cost of wet limestone SO_2 control unit	M\$/year
BARC	The bare cost of the control unit	M\$

AOC	Annual net operating cost	M$
TCR	Total capital required	M$
CONTIN	Contingency	
F	Location Factor	

TABLE 5.2

UNIT COSTS USED IN ILLUSTRATIVE EXAMPLES - WET LIMESTONE STACK GAS SCRUBBING MODEL

Purchased Price of Limestone (\$/Ton)	4.00
Purchased Price of Ammonia (\$/Ton)	50.00
Purchased Price of Water (\$/MGal)	0.20
Purchased Price of Fuel Oil (\$/MMBtu)	0.80
Purchased Price of Electricity (mils/Kwhr)	8.00
Average Hourly Wages Per Gulf Coast (\$/Hr)	7.00
Interest on Capital During Construction (%)	12.00

TABLE 5.3

WET LIMESTONE PROCESS AND COST MODEL
SUMMARY OF EQUATIONS

$$AL = 600\ CL \cdot LF\ (SF/28) \qquad \text{M\$/year}$$

$$AA = 0.43\ CA\ (SF/28) \qquad \text{M\$/year}$$

$$AW = 230\ CW \cdot LF\ [\ (GP/3{,}300) + (SF/28)] \qquad \text{M\$/year}$$

$$AF = 1{,}800\ CF \cdot LF\ (GP/3{,}300) \qquad \text{M\$/year}$$

$$AE = CE \cdot LF\ [213\ (GP/3{,}300) + 35\ (SF/28)] \qquad \text{M\$/year}$$

$$ANR = AL + AA + AW + AF + AE \qquad \text{M\$/year}$$

$$EC = \sum_{n=1}^{NA} RB\ [1041\ (GT/550)^{0.5} + 408\ (GT/550)^{0.9}]_n \qquad \text{M\$}$$

$$+238\ RP\ (GP/3{,}300)^{0.5} + 201\ (SF/28)^{0.5}$$

$$ES = 1680\ (SF/28)^{0.9} \qquad \text{M\$}$$

$$P = 5{,}000 \left[\frac{SF \cdot LF}{28}\right]^{0.9} \qquad \text{M\$}$$

$$BARC = 1.15\ (E + M) + (P + 1.43\ L)F \qquad \text{M\$}$$

$$TPI = 1.12\ (1.0 + CONTIN)\ BARC \qquad \text{M\$}$$

$$TAC = 0.237\ TPI + 2.1\ TO \cdot CO\ (1+F) + 1.04\ ANR \qquad \text{M\$/year}$$

$$TCR = 1.15\ TPI + 0.8\ TO \cdot CO\ (1+F) + 0.4\ ANR \qquad \text{M\$}$$

FIGURE 5.1 WET LIMESTONE PROCESS FLOWSHEET

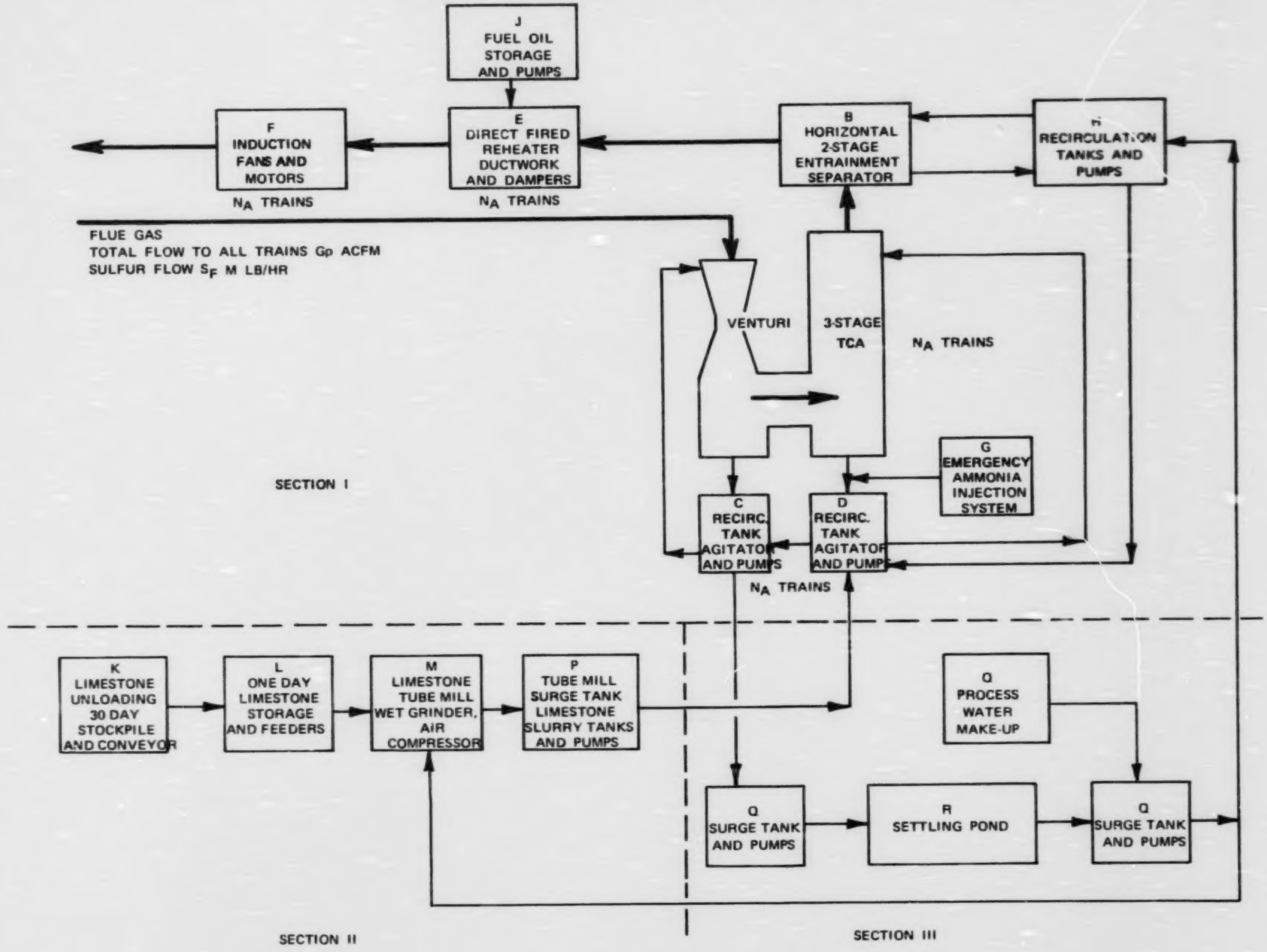
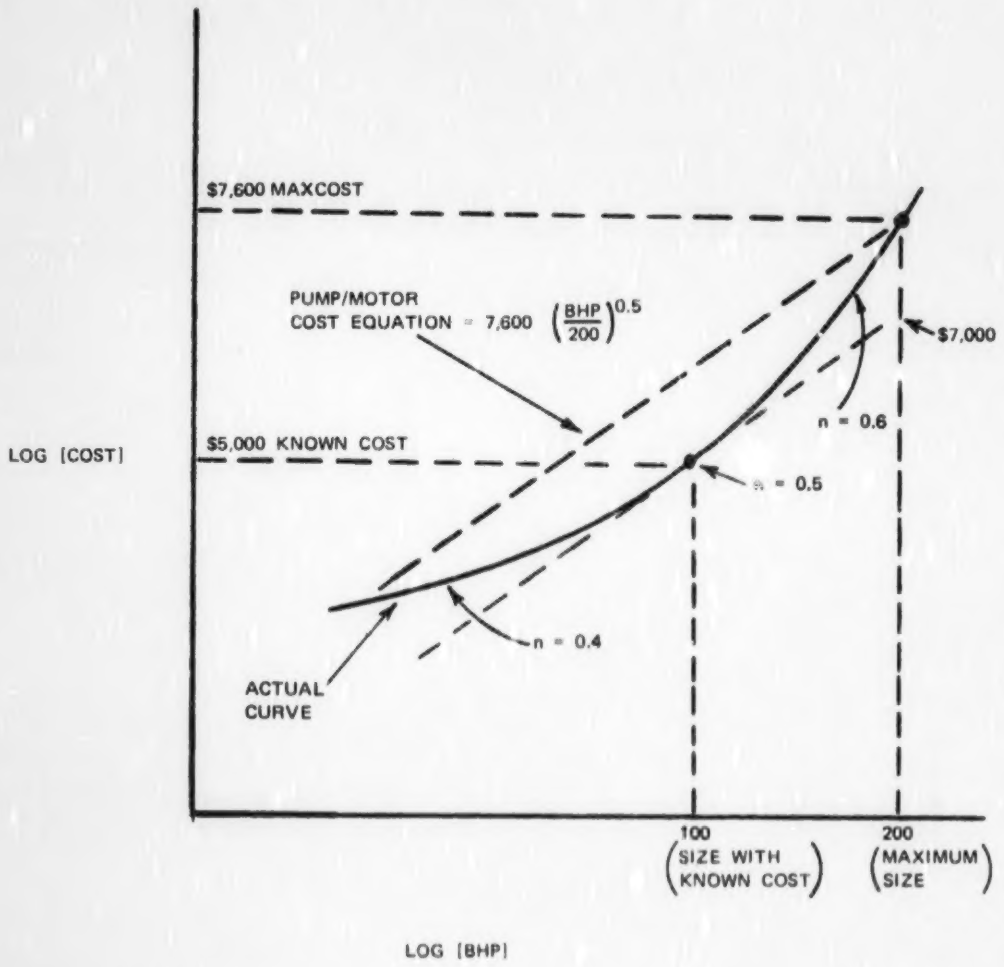

J
FUEL OIL STORAGE AND PUMPS
F
INDUCTION FANS AND MOTORS
N_A TRAINS
E
DIRECT FIRED REHEATER DUCTWORK AND DAMPERS
N_A TRAINS
B
HORIZONTAL 2-STAGE ENTRAINMENT SEPARATOR
H
RECIRCULATION TANKS AND PUMPS
FLUE GAS
TOTAL FLOW TO ALL TRAINS G_P ACFM
SULFUR FLOW S_F M LB/HR
VENTURI
3-STAGE TCA
N_A TRAINS
G
EMERGENCY AMMONIA INJECTION SYSTEM
C
RECIRC. TANK AGITATOR AND PUMPS
D
RECIRC. TANK AGITATOR AND PUMPS
N_A TRAINS
SECTION I
K
LIMESTONE UNLOADING 30 DAY STOCKPILE AND CONVEYOR
L
ONE DAY LIMESTONE STORAGE AND FEEDERS
M
LIMESTONE TUBE MILL WET GRINDER, AIR COMPRESSOR
P
TUBE MILL SURGE TANK LIMESTONE SLURRY TANKS AND PUMPS
Q
PROCESS WATER MAKE-UP
Q
SURGE TANK AND PUMPS
R
SETTLING POND
Q
SURGE TANK AND PUMPS
SECTION II
SECTION III

FIGURE 5.2

METHOD OF VARYING EQUIPMENT COST WITH SIZE

PUMP/MOTOR COST = $K(\text{BHP})^n$



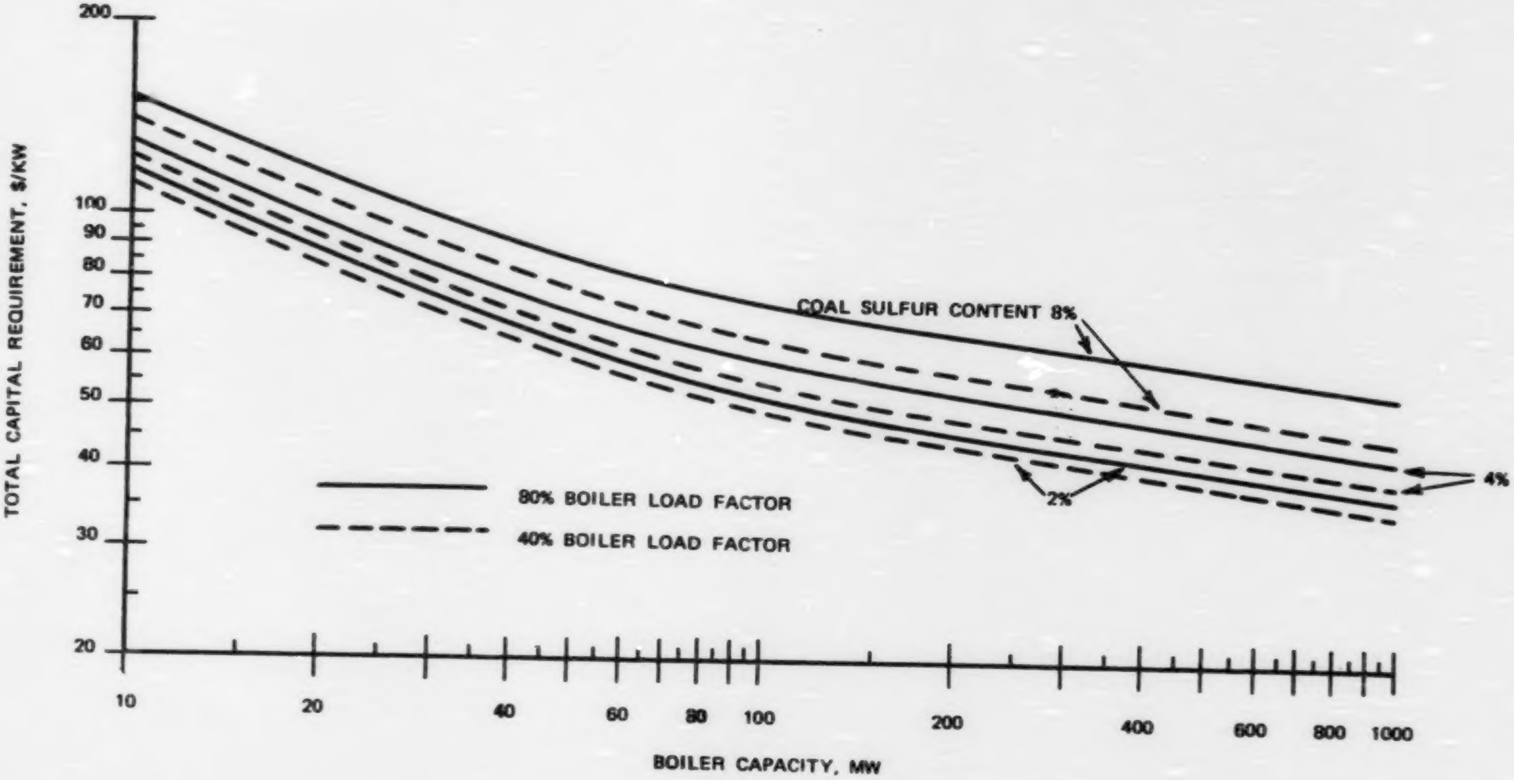
MAXCOST = $5000\left(\frac{200}{100}\right)^{0.6}$ = $7,600

COST EQUATION

PUMP/MOTOR COST = $7{,}600\left(\frac{\text{BHP}}{200}\right)^{0.5}$

FIGURE 5.3

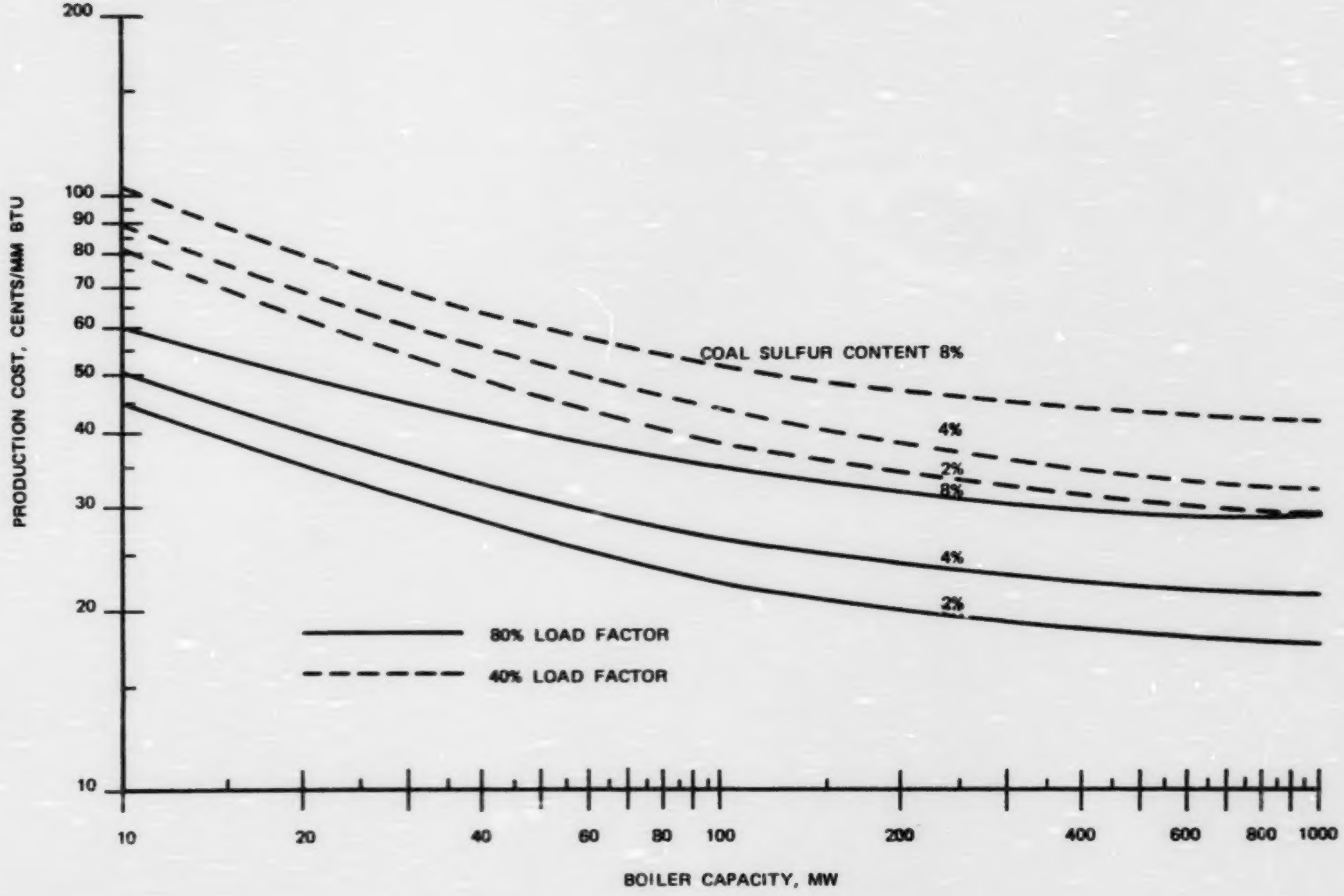
EFFECT OF BOILER CAPACITY ON TOTAL CAPITAL REQUIREMENT
WET LIMESTONE PROCESS



BASIS OF CALCULATION = NO RETROFIT, NO CONTINGENCY, U.S. GULF COAST LOCATION, END OF 1973 FIGURE, BOILER HEAT RATE 9,500 BTU/KWH, HEATING VALUE OF COAL (HHV) OF 11,000 BTU/LB.

FIGURE 5.4

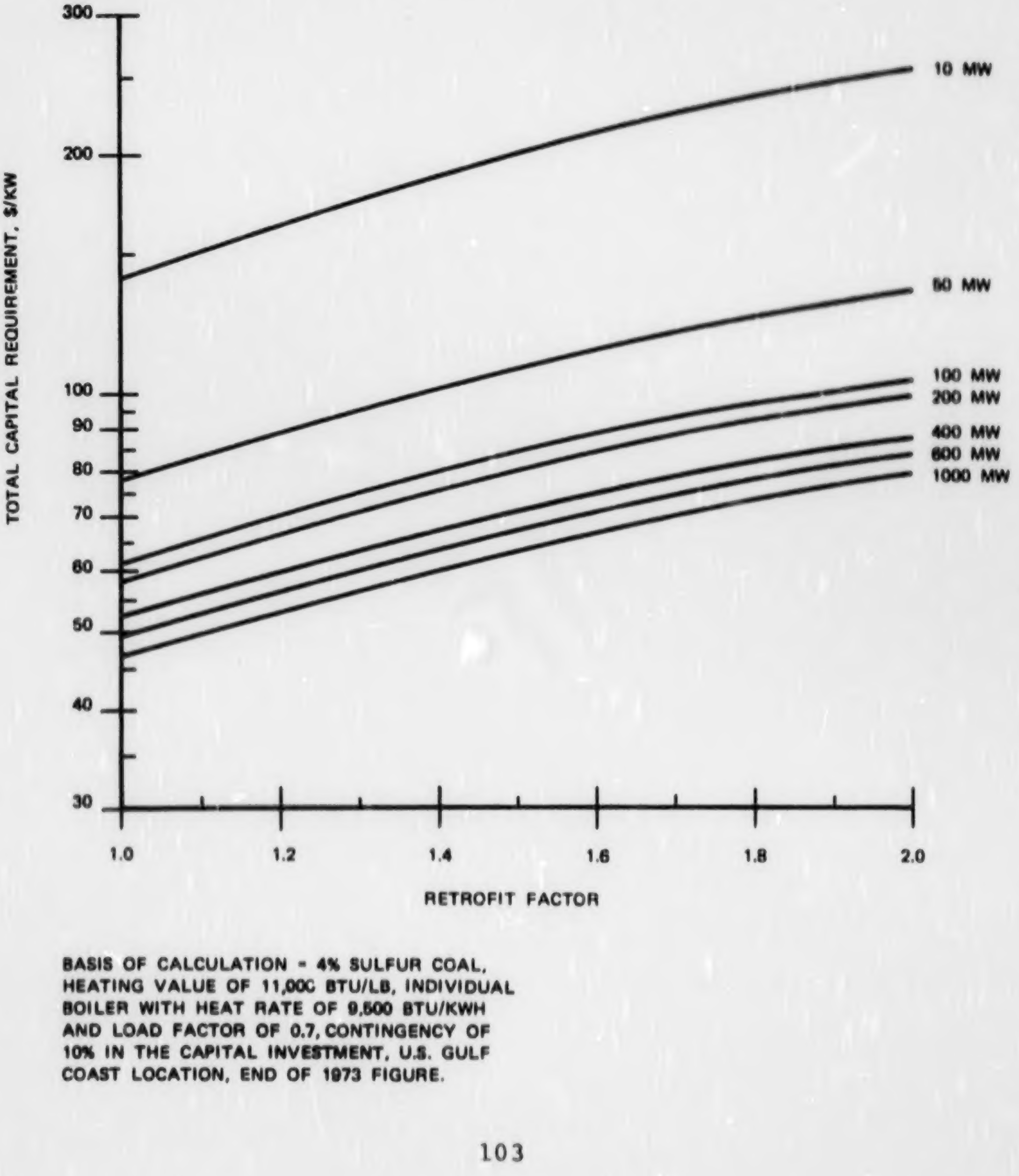
EFFECT OF BOILER CAPACITY ON PRODUCTION COST
WET LIMESTONE PROCESS



(SEE FIGURE 5.3 FOR BASIS OF CALCULATION.)

FIGURE 5.5

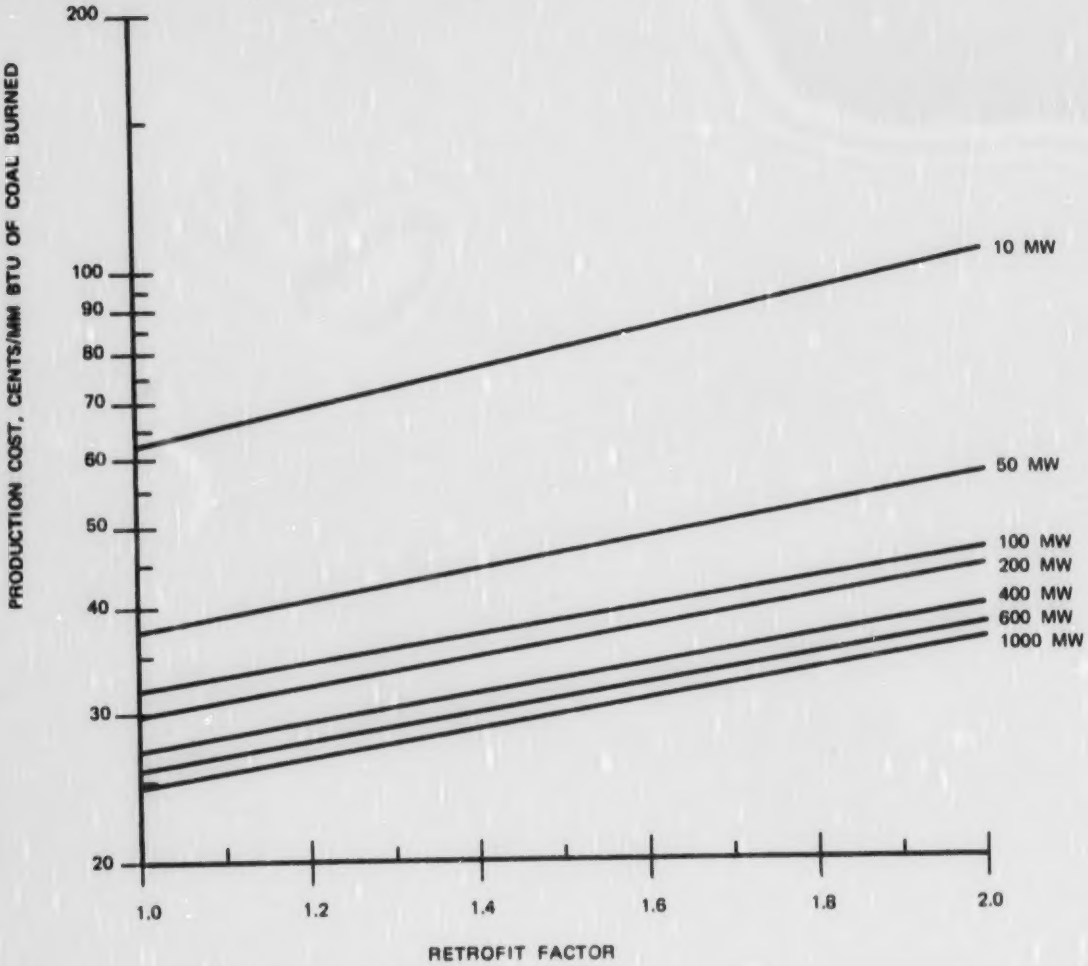
EFFECT OF BOILER RETROFIT DIFFICULTY ON TOTAL CAPITAL REQUIREMENT WET LIMESTONE PROCESS



BASIS OF CALCULATION - 4% SULFUR COAL, HEATING VALUE OF 11,000 BTU/LB, INDIVIDUAL BOILER WITH HEAT RATE OF 9,500 BTU/KWH AND LOAD FACTOR OF 0.7, CONTINGENCY OF 10% IN THE CAPITAL INVESTMENT, U.S. GULF COAST LOCATION, END OF 1973 FIGURE.

FIGURE 5.6

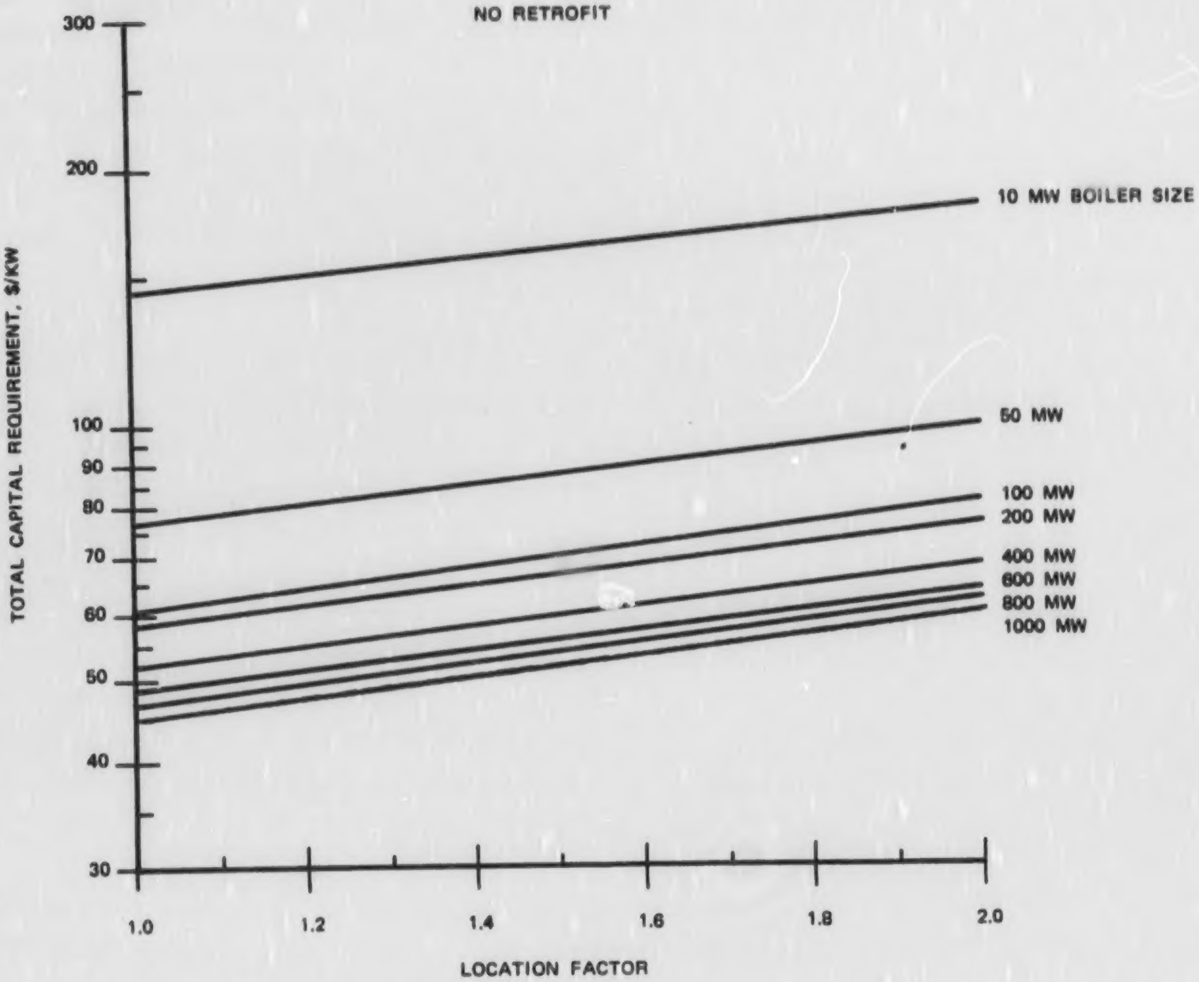
EFFECT OF BOILER RETROFIT DIFFICULTY ON PRODUCTION COST WET LIMESTONE PROCESS



(SEE FIGURE 5.5 FOR BASIS OF CALCULATION.)

FIGURE 5.7

EFFECT OF LOCATION FACTOR ON TOTAL CAPITAL REQUIREMENT WET LIMESTONE PROCESS



BASIS OF CALCULATION: 4% SULFUR COAL, HEATING VALUE 11,000 BTU/LB, INDIVIDUAL BOILER WITH HEAT RATE OF 9,500 BTU/KWH AND LOAD FACTOR OF 0.7, CONTINGENCY 10% IN CAPITAL INVESTMENT, U.S. GULF COAST LOCATION, END OF 1973 FIGURE.

6. THE WELLMAN/ALLIED PROCESS

6.1 Process Appraisal

The basis for the Wellman/Allied process and cost models is the design proposed for the demonstration plant to be installed at the D.H. Mitchell plant of the Northern Indiana Public Service Company. This system is a combination of the Wellman-Lord SO_2 recovery process and the Allied Chemical SO_2 reduction process, producing elemental sulfur as an end product. The process and cost models were developed for this combined system. For simplicity, it will hereafter be referred to in this report as the Wellman/Allied system or process.

The NIPSCO design, which is for a 115 MW plant burning 3.2% sulfur coal, is the only one available which is sufficiently detailed for use in deriving the models. While a review of the design showed it to be reasonable, a number of process changes were made and are discussed below.

The NIPSCO design provides for an absorber capable of handling the maximum flow of flue gas from the power plant whereas the SO_2 recovery system is designed for the average flow rate, corresponding to an 80% load factor. The difference between these design capacities is handled by providing large surge capacity for the sodium sulfite solution. For the model, this has been simplified to allow the entire regeneration plant to run at full absorber capacity, i.e., at 100% load factor. Correspondingly, surge capacity has been reduced. In addition to being simpler to model (i.e., does not require knowledge of short-term variation of boiler load factor), this type of design provides the capability of operating the regeneration and recovery system at peak load conditions if it were necessary, thus giving the total system a greater flexibility and operating range. Under normal operation, the regeneration and recovery system would operate at some reduced steady state level, while allowing the scrubbing section to fluctuate in response to varying demands from the boiler. The reduced level of operation would have to be adjusted periodically, depending on solution inventory and anticipated boiler operation. Of course, this type of design would be somewhat more expensive than designing for the average flow (see section 6.6).

The NIPSCO absorber was designed to remove about 91% of the SO_2 in the flue gas and included three contacting stages plus space for a fourth. In lieu of operating data, it was assumed that four trays would be a safe design for 90% removal. An additional tray was added for the model and it was assumed that this would give an overall SO_2 removal efficiency of 95%. This high removal efficiency was chosen as the "standard" design to permit use of the model not only in utility applications, but also in other applications (smelters, Claus plants, etc.) where high SO_2 removal efficiencies would be particularly desirable. However, high removal efficiency is also a useful device for investigating utility plant applications, since the computer cost program has been designed to consider sequential control of plant boilers until the desired emissions limitations are achieved (see section 7.1).

The flue gas blower has been changed from upstream of the absorber to downstream of the reheater. Pressure drop has been increased from 18" H_2O to 30" H_2O and is distributed as follows:

	Inches H_2O
Prescrubber	6
5 Absorption Stages	15
Demister	2
Ductwork	7
	30

Some items which are identical in design and operation to those which were included in the wet limestone model previously developed have been based on that model rather than the NIPSCO design. These are: the flue gas ductwork and dampers, the gas reheater, and the reheat fuel storage and delivery system.

All pumps, fans, blowers, and compressors have been spared with the exception of the flue gas fan and the SO_2 compressor. In cases where multiple units are used for a single service, it was assumed that one common spare would be adequate.

The NIPSCO design included a separate stack for the clean flue gas. This was deleted from the model since the clean flue gas would ordinarily be diverted back to the main stack, which is properly costed to the source plant and not the control unit.

The process flow sheet is shown in Figure 1 and is divided into four main areas:

1) The absorber area, including gas reheat and compression
2) The SO_2 regeneration area
3) The purge/make-up area
4) The SO_2 reduction area (Allied Process)

6.2 Evaluation of the NIPSCO Project Cost Estimate

The cost estimate made by Davy Powergas and Allied Chemical for the NIPSCO project has been used as a basis for the cost model. Since process equipment costs for this project were more than 85% quoted, they should form a sound basis for defining equipment costs in the model. Quotes were received primarily during the latter part of 1972 and have been assumed to be valid as of the end of 1972. Before using these costs, they have been increased by 5% to allow for escalation to the end of 1973, which is the reference time chosen for the model.

For some pieces of equipment in the absorber area that are common in design and operation to both the Wellman/Allied system and the wet limestone process, costs have been derived from the wet limestone model rather than the NIPSCO estimate. These are:

1. Induction fan
2. Reheater, ductwork, and dampers
3. Fuel oil system

6.3 Variation of Equipment Costs with Plant Size

In order to determine exponents relating cost to size for different types of equipment, several sources were consulted (9, 14, 21). This resulted in the following variation of equipment cost with size for the different types of equipment used in the Wellman/Allied system:

	Cost Proportional to
Tower shells (including lining)	$(\text{ACFM})^{0.4-0.6}$
Tower internals	$(\text{ACFM})^{0.8-1.0}$
Centrifugal pumps	$(\text{BHP})^{0.4-0.6}$
Tanks and drums	$(\text{volume})^{0.4-0.6}$
Agitators	$(\text{BHP})^{0.4-0.6}$
Pressure filter	$(\text{flow})^{0.5-0.7}$
Fans, blowers, and compressors	$(\text{BHP})^{0.8-1.0}$
Direct-fired heaters	$(\text{duty})^{0.4-0.6}$
Ductwork and dampers	$(\text{ACFM})^{0.4-0.6}$
Heat exchangers	$(\text{surface})^{0.5-0.7}$
Forced-circulation evaporators (complete system)	$(\text{duty})^{0.5-0.7}$
Storage silos and bins	$(\text{volume})^{0.8-1.0}$
Entrainment separators	$(\text{ACFM})^{0.8-1.0}$
Pressure vessels	$(\text{volume})^{0.4-0.6}$
Pressure vessel internals	$(\text{ACFM})^{0.8-1.0}$
Sulfur pit	$(\text{volume})^{0.8-1.0}$
Miscellaneous solids handling equipment	$(\text{flow})^{0.8-1.0}$

Equipment sizes for the process were related to either of two basic variables: the flue gas flow or the flow of sulfur in the flue gas. The absorber and related equipment, which are all included in the absorber area, are proportional in size to the flue gas flow. The remainder of the process equipment in the plant is proportional to sulfur flow. Each process section was reviewed to determine its maximum train size, based on the equipment sizes shown for the NIPSCO design.

An analysis of the absorber area showed that the maximum size absorber can handle a gas flow of about 550,000 ACFM. The other gas-related equipment has also been limited to this maximum size, with the exception of the fuel oil system. The latter, which was taken from the wet limestone model, has been assumed to be single train, regardless of plant size. For the sulfur-related equipment in the absorber area, an upper limit on equipment sizes was found at a train size corresponding to about 7,000 lbs/hr of sulfur. This was also the case for the SO_2 regeneration area. Equipment in the purge/make-up area can be single train at the reference plant flow of 28,000 lbs/hr of sulfur, but this is about the maximum practical size for this section. Within the size ranges of interest, no upper limit was found for single train operation in the SO_2 reduction area (Allied plant).

Considering the reference plant size (3,300 MACFM of flue gas and 28,000 lbs/hr of sulfur), and based on the NIPSCO design, scale-up factors to maximum or reference size trains were determined to be:

Area	Scale-up Factor	Number of Trains in Reference Size Plant
Absorber Area		
gas-related equipment	1.18	6
sulfur-related equipment	3.04	4
SO_2 Regeneration Area	3.04	4
Purge/Make-up Area	12.16	1
SO_2 Reduction Area (Allied plant)	12.16	1

6.4 Cost Model

6.4.1 Equipment Costs

Using the NIPSCO estimate as a basis and the sacle-up factors given in the preceding section, equipment costs have been calculated for the maximum or reference size units and are shown below.

1. The Absorber Area

	Cost of Maximum Size Train M$ (end of 73)	Cost Relationship
A. Absorber shell, lining and circulation pumps[1]; prescrubber circulation pumps; reheater, ductwork and dampers	726	$GT^{0.5}$
B. Vessel internals; induction fan	639	$GT^{0.9}$
C. Fuel oil system[2]	119	$GP^{0.9}$
D. Tanks[3], pumps, and agitators	113	$S7^{0.5}$
E. Fly ash filter system	127	$S7^{0.6}$

where GT is the flue gas flow per train in MACFM, GP is the total plant flue gas flow in MACFM, and S7 is the sulfur flow rate per train in Mlbs/hr.

For the reference size plant, six maximum size absorber trains, one fuel oil system, and four maximum size sulfur trains are required. Thus, the total equipment cost for the absorber area, EA, is:

---

[1] NIPSCO costs have been adjusted to provide for an additional absorption stage.

[2] Unit cost taken from wet limestone model.

[3] NIPSCO costs have been adjusted to reflect reduced surge capacity.

$$
\begin{aligned}
EA &= (726 + 639) \times (6 \text{ trains}) \\
&\quad + (119) \times (1 \text{ train}) \\
&\quad + (133 + 127) \times (4 \text{ trains}) \\
&= \$9394M
\end{aligned}
$$

In general, for a plant with a total gas flow of GP MACFM, a gas flow per train of GT MACFM (with NA absorber trains), a total sulfur rate of SF M lbs/hr, and a sulfur rate per train of S7 M lbs/hr (with N7 sulfur trains), the total equipment cost for the absorber area is:

$$
EA = \sum_{n=1}^{NA} \left[726\ (GT/550)^{0.5} + 639\ (GT/550)^{0.9} \right]_n + 119\ (GP/3300)^{0.5} + \left[133\ (S7/7)^{0.5} + 127\ IF\ (S7/7)^{0.6} \right] N7 \qquad M\$
$$

where

$S7 = SF/N7$

$N7 = SF/7$ (rounded to next higher integer)

IF is merely an index used to include or delete the cost of a fly ash filter system, as necessary.

$IF = 1$ if particulates are present in the flue gas

$IF = 0$ if particulates are absent from the flue gas

2. The SO_2 Regeneration Area

	Cost of Maximum Size Train, M$ (end of 73)	Cost Relationship
A. Vessels, agitators and pumps[1]	209	$S7^{0.5}$
B. Heat exchangers, evaporator system	618	$S7^{0.6}$
C. Compressor, vessel internals	157	$S7^{0.9}$

For the reference size plant, four trains are required, each handling the maximum sulfur rate per train of 7 M lbs/hr. The total equipment cost for the SO_2 regeneration area is:

$$\begin{aligned} ES &= (209 + 618 + 157) \times (4 \text{ trains}) \\ &= \$3936M \end{aligned}$$

In general, for a plant handling a sulfur rate per train of S7 M lbs/hr (with N7 sulfur trains), the total equipment cost for the SO_2 regeneration area is:

$$ES = \left[209\ (S7/7)^{0.5} + 618\ (S7/7)^{0.6} + 157\ (S7/7)^{0.9}\right] N7 \quad M\$$$

3. The Purge/Make-up Area

	Cost of Maximum Size Train, M$ (end of 73)	Cost Relationship
A. Pumps, tanks, agitators, heat exchangers, and dryer	525	$S28^{0.5}$
B. Separating Equipment	380	$S28^{0.6}$

---

[1] NIPSCO costs for absorber feed tank and agitator have been adjusted to reflect reduced surge capacity.

	Cost of Maximum Size Train, M$ (end of 73)	Cost Relationship
C. Special equipment	86	$S28^{0.7}$
D. Packaged heat exchanger	306	$S28^{0.8}$
E. Fan and miscellaneous solids handling equipment	519	$S28^{0.9}$

For the reference size plant, one train is required, handling the maximum sulfur rate per train of 28 M lbs/hr. The total equipment cost for the purge/make-up area is:

$$\begin{aligned} EP &= (525 + 380 + 86 + 306 + 519) \times (1 \text{ train}) \\ &= \$1816M \end{aligned}$$

In general, for a plant handling a sulfur rate per train of S28 M lbs/hr (with N28 sulfur trains), the total equipment cost for the purge/make-up area is:

$$\begin{aligned} EP = \Big[& 525\ (S28/28)^{0.5} + 380\ (S28/28)^{0.6} + 86\ (S28/28)^{0.7} \\ & + 306\ (S28/28)^{0.8} + 519\ (S28/28)^{0.9} \Big] N28 \qquad M\$ \end{aligned}$$

where N28 = SF/28 (rounded to next higher integer)
S28 = SF/N28

4. The SO_2 Reduction Area (Allied plant)

	Cost of Reference Size Train, M$ (end of 73)	Cost Relationship
A. Pumps, fired heaters, vessels, and ductwork	998	$SF^{0.5}$
B. Heat exchangers	287	$SF^{0.6}$
C. Compressors, mist eliminator, sulfur pit, and vessel internals	683	$SF^{0.9}$

For the reference size plant, this area is sized for an equivalent sulfur rate of 28 M lbs/hr. The total equipment cost for the SO_2 reduction area is:

$$\begin{aligned} ER &= 998 + 287 + 683 \\ &= \$1968M \end{aligned}$$

In general, for a plant handling a sulfur rate of SF M lbs/hr the total equipment cost for the SO_2 reduction area is:

$$ER = 998\ (SF/28)^{0.5} + 287\ (SF/28)^{0.6} + 683\ (SF/28)^{0.9} \quad M\$$$

The total equipment costs for the Wellman/Allied system can be summarized as follows:

$$EA = \sum_{n=1}^{NA} RB\ \left[726\ (GT/550)^{0.5} + 369\ (ST/550)^{0.9}\right]_n$$

$$+ 119\ RP\ (GP/3300)^{0.5} + \left[133\ (S7/7)^{0.5} + 127\ IF\ (S7/7)^{0.6}\right] N7 \quad M\$$$

$$ES = \left[209\ (S7/7)^{0.5} + 618\ (S7/7)^{0.6} + 157\ (S7/7)^{0.9}\right] N7 \quad M\$$$

$$EP = \left[525\ (S28/28)^{0.5} + 380\ (S28/28)^{0.6} + 86\ (S28/28)^{0.7}\right.$$

$$\left. + 306\ (S28/28)^{0.8} + 519\ (S28/28)^{0.9}\right] N28 \quad M\$$$

$$ER = 998\ (SF/28)^{0.5} + 287\ (SF/28)^{0.6} + 683\ (SF/28)^{0.9} \quad M\$$$

RB and RP are the retrofit difficulty factors as described in the wet limestone process model.

6.4.2 Other Material Costs and Labor Costs

Costs for labor and other materials generally can be estimated as a percentage of major equipment costs. Since the NIPSCO estimate was broken down by plant area, factors were obtained for these costs for each area. The factors derived from the data are shown below, where E is the major equipment cost, L is the labor cost, and M is the cost of other materials. The letters A, S, P, and R refer to the absorber area, the SO_2 regeneration area, the purge/make-up area, and the SO_2 reduction area respectively. Labor costs are based on the Gulf Coast area. Field materials include only piping, instruments, electrical, insulation, painting, concrete, and structural steel.

$$L_A = 0.224\ E_A \qquad M_A = 0.429\ E_A$$
$$L_S = 0.310\ E_S \qquad M_S = 0.742\ E_S$$
$$L_P = 0.433\ E_P \qquad M_P = 0.827\ E_P$$
$$L_R = 0.623\ E_R \qquad M_R = 0.772\ E_R$$

6.4.3 Raw Materials and Utilities Costs

1. Sodium Carbonate

Sodium carbonate make-up is required to replenish the sodium values lost by oxidation of the scrubbing solution. The quantity used is directly proportional to the sulfur rate, SF. For the NIPSCO design, 0.265 tons/hr were required. Since for the reference size plant the scale-up factor on the sulfur rate is 12.16, the sodium carbonate make-up for the reference plant at 100% load factor is:

$$\text{Consumption} = \frac{0.265}{1000} \times 12.16 \times 8760 = 28.2 \text{ M tons/yr}$$

In general, the annual cost of sodium carbonate, AS, for a power plant having a load factor of LF is:

$$AS = 28.2\ CS \cdot LF\ (SF/28) \qquad \text{M\$/yr}$$

where CS is the purchase price of sodium carbonate in $/ton.

2. Natural Gas

Natural gas is used in the SO_2 reduction area to convert the SO_2 to elemental sulfur. The amount consumed is proportional to the sulfur rate, and for the NIPSCO plant equals 13.7 MSCFH. The annual consumption for the reference plant at 100% load factor is:

$$\text{Consumption} = \frac{13.7}{1000} \times 12.16 \times 8760 = 1460 \text{ MMSCF/yr}$$

The annual cost of natural gas, AN, is:

$$AN = 1460\ CN \cdot LF\ (SF/28) \qquad \text{M\$/yr}$$

where CN is the purchase price of natural gas in $/MSCF.

3. Filter Aid

Filter aid, which is used in the fly ash filter system, is of course needed only if particulates are present in the flue gas. The quantity required is assumed to be proportional to the gas flow. A design rate of 40 lbs/hr was shown for the NIPSCO design. The scale-up factor to the reference size plant, on gas flow, is 6 x 1.18, or 7.08. For the reference plant, therefore, the annual consumption at 100% load factor is:

$$\text{Consumption} = \frac{40}{2000} \times \frac{7.08}{1000} \times 8760 = 1.24 \text{ M tons/yr}$$

The annual cost of filter aid, AFA, is:

$$AFA = 1.24\ CFA \cdot LF \cdot IF\ (GP/3300) \qquad \text{M\$/yr}$$

where CFA is the purchase price of filter aid in \$/ton and IF is the fly ash index previously defined.

4. Power

The power consumption shown for the NIPSCO design has been adjusted to reflect some process and equipment changes (an additional absorption stage, increased gas pressure drop, etc.) which were incorporated in the model, as discussed previously. The adjusted power requirement for the NIPSCO design is 3220 KW of which 2480 KW are proportional to the gas flow rate and 740 KW are proportional to the sulfur rate.

The annual power consumption of the reference plant at 100% load factor is:

$$\begin{aligned} \text{Consumption} &= 2.480 \times 7.08 \times 8760 \text{ (proportional to GP)} \\ &+ 0.740 \times 12.16 \times 8760 \text{ (proportional to SF)} \\ &= 154{,}000 \text{ MKWH/hr.} + 79{,}000 \text{ MKWH/yr} \end{aligned}$$

The annual power cost, AE, is:

$$AE = \left[154\ (GP/3300) + 79\ (SF/28)\right] CE \cdot LF \qquad \text{M\$/yr}$$

where CE is the purchase (or transfer) price of electricity in mills/KWH.

5. Steam

The steam consumption shown for the NIPSCO design has been adjusted because of the deletion of steam turbine drives on the flue gas fan and the SO_2 compressor. The adjusted value is 51.0 M lbs/hr. and is proportional to the sulfur rate. For the reference plant at 100% load factor, the steam consumption is:

$$\text{Consumption} = \frac{51.0}{1000} \times 12.16 \times 8760 = 5430 \text{ MM lbs/yr.}$$

The annual cost of steam, AH, is:

$$AH = 5430\ CH \cdot LF\ (SF/28) \qquad \text{M\$/yr}$$

where CH is the purchase (or transfer) price of steam in $/M lbs.

6. Cooling Water

The total cooling water requirement for the NIPSCO plant is 3.34 MGPM of which 0.23 MGPM is proportional to the gas flow and 3.11 MGPM is proportional to the sulfur rate. Cooling water required for the reference plant at 100% load factor is:

$$\begin{aligned}\text{Consumption} &= 0.23 \times 7.08 \times 60 \times 8760 \text{ (proportional to GP)}\\ &+ 3.11 \times 12.16 \times 50 \times 8760 \text{ (proportional to SF)}\\ &= 856{,}000 \text{ M gal/yr} + 19{,}900{,}000 \text{ M gal/yr}\end{aligned}$$

The annual cost of cooling water, ACW, is:

$$ACW = \left[856\ (GP/3300) + 19{,}900\ (SF/28)\right] CCW \cdot LF \qquad \text{M\$/yr}$$

where CCW is the cost of cooling water in $/M gal.

7. Process Water

Small amounts of process water are used in the purge and make-up systems and are proportional to sulfur rate. For NIPSCO, process water use is about 10 GPM. The quantity required for the reference plant at 100% load factor is:

$$\text{Consumption} = \frac{10}{1000} \times 12.16 \times 60 \times 8760 = 64{,}000 \text{ M gal/yr}$$

The annual cost of process water, AW, is:

$$AW = 64\ (SF/28)\ CW \cdot LF \qquad \text{M\$/yr}$$

where CW is the cost of process water in $/M gal.

8. Fuel Oil

Since the fuel oil system for reheating the flue gas is identical to that included in the wet limestone model, the oil consumption and cost will be the same. For the reference plant at 100% load factor:

Consumption = 1,800,000 MM Btu/yr.

The cost of fuel oil, AF, is:

$$AF = 1{,}800\ (GP/3300)\ CF \cdot LF \qquad \text{M\$/yr}$$

where CF is the purchase price of fuel oil in $/MM Btu.

9. Credits

The process produces two materials: sulfur, and a dry purge solids stream consisting of sodium sulfite, sodium sulfate, and and sodium thiosulfate. The product sulfur would normally be listed as a credit. However, the purge solids may have positive or negative value depending upon whether or not they are salable. Normally, it is expected that a waste disposal cost would be incurred. The cost treatment of the purge solids can be handled by insertion of a positive or negative unit value in the model.

a. Sulfur

The sulfur production for the NIPSCO plant is 21.5 long tons/day and is proportional to the sulfur rate. For the reference plant at 100% load factor:

$$\text{Production} = \frac{21.5}{1000} \times 12.16 \times 365 = 95.4 \text{ M long tons/yr}$$

The sulfur credit, ASC, is:

$$ASC = 95.4\ (SF/28)\ VSC \cdot LF \qquad M\$/yr$$

where VSC is the unit value of sulfur in $/long ton.

b. Purge Solids

The NIPSCO design shows a purge solids production rate of 0.35 tons/hr which is proportional to the sulfur rate. The purge solids flow for the reference plant at 100% load factor is:

$$\text{Production} = \frac{0.35}{1000} \times 12.16 \times 8760 = 37.3 \text{ M tons/yr}$$

The purge solids credit (or debit), APS, is:

$$APS = 37.3\ (SF/28)\ VPS \cdot LF \qquad M\$/yr$$

where VPS is the unit value of the purge solids in $/ton. If the purge solids are listed as a credit (debit), VPS would be positive (negative).

The total cost of raw materials and utilities less credits, ANR, is:

$$ANR = AS + AN + AFA + AE + AH + ACW + AW + AF - ASC - APS \quad M\$/yr$$

6.4.4 Total Plant Investment and Total Capital Required

The bare cost (BARC), total plant investment (TPI), and total capital required (TCR) for the Wellman/Allied system can be calculated from the appropriate equations in the General Cost Model. Thus,

$$\begin{array}{llll} & \mathrm{BARC} & = 1.15\ (\mathrm{E+M}) + 1.43\ \mathrm{L \cdot F} & \mathrm{M\$} \\ & \mathrm{TPI} & = 1.12\ (1.0 + \mathrm{CONTIN}) \cdot \mathrm{BARC} & \mathrm{M\$} \\ & \mathrm{TCR} & = 1.15\ \mathrm{TPI} + 0.8\ \mathrm{TO \cdot CO}\ (1.0+\mathrm{F}) + 0.4\ \mathrm{ANR} & \mathrm{M\$} \\ \text{where} & \mathrm{E} & = \mathrm{EA + ES + EP + ER} & \mathrm{M\$} \\ & \mathrm{M} & = \mathrm{MA + MS + MP + MR} & \mathrm{M\$} \\ & \mathrm{L} & = \mathrm{LA + LS + LP + LR} & \mathrm{M\$} \end{array}$$

6.4.5 Operating Costs

The total net annual operating cost, AOC, represents the total cost of running the plant, excluding depreciation, interest, and income tax. It is given by the following equation from the General Cost Model:

$$\mathrm{AOC} = 0.078\ \mathrm{TPI} + 2.0\ \mathrm{TO \cdot CO}\ (1.0 + \mathrm{F}) + \mathrm{ANR} \qquad \mathrm{M\$/yr}$$

where TO = total number of shift operators
CO = hourly rate of operators

For plants larger than 200 MW, the Wellman/Allied process requires 16 operators (4 per shift). It has been assumed that for plants less than 200 MW, operating labor costs are directly proportional to plant size.

The total annual production cost, TAC, including the return on capital, interest, and income tax is given by:

$$\mathrm{TAC} = 0.237\ \mathrm{TPI} + 2.1\ \mathrm{TO \cdot CO}\ (1.0 + \mathrm{F}) + 1.04\ \mathrm{ANR} \qquad \mathrm{M\$/yr}$$

6.5 Effect of Various Parameters on Costs

Figures 6.2-6.6 show typical costs which were calculated from the model illustrating the effects of different variables on plant costs. Unit values used for raw materials and utilities are listed in Table 6.1.

In general, the effects of variables are similar to those noted for the Wet Limestone process. However, there are some important differences. Costs for Wellman/Allied are much greater than Wet Limestone for small boilers and high percent sulfur. For all sizes, percent sulfur has a greater impact on costs in the Wellman/Allied system.

Load factor has a significant effect on operating costs, as in the Wet Limestone process, particularly for small sizes. For large plants, percent sulfur has a greater effect than load factor.

Retrofit factor is less significant for the Wellman/Allied process compared to Wet Limestone. This is due to the fact that much of the cost is in sulfur recovery rather than scrubbing. The latter is usually where the retrofit difficulty occurs.

6.6 Wellman/Allied Process Variations and Impact on Costs

For convenience and simplicity, the model developed for the Wellman/Allied system uses a single processing scheme. There are, however, process modifications which could be made and which could effect costs.

The model assumes that the SO_2 recovery and reduction sections would be designed for full capacity. Designing for less than full capacity is possible, as was done for the NIPSCO project. This requires some accurate knowledge of the expected variation of load factor with time. If this information is available and indicates that a less than full capacity design is possible, the model could easily be changed to accomodate this. The sulfur flow in the gas would be reduced appropriately before using the equipment cost equations. For example, if the recovery sections were to be designed for 80% of full capacity, the design sulfur rate would be:

$$S_F' = 0.8\ S_F$$

S_F' would then be used to cost equipment. Calculations with the model show that sizing recovery sections for 80% of capacity would reduce capital costs by about 6-8% below costs for a 100% design.

The model uses a single effect evaporator (for SO_2 regeneration) and assumes electric drive for the flue gas fans. Steam economies could be achieved by double-effect evaporation and steam drive for the fans. Low pressure exhaust steam from the fans would be used in the first effect of the evaporator. By this method, overall energy consumption of the process could be reduced from more than 11% of the power plant heat input to perhaps 8-9%. However, it is likely that capital costs would increase. In addition, a double-effect evaporator with its first effect operating at a higher temperature suggests the likelihood of increased sulfite oxidation losses. This would increase the make-up cost. Although there were not enough data available to estimate costs for this type of design, it appears unlikely that costs could be reduced significantly.

It has been assumed for all models that costs of multiple trains are direct multiples of single train costs. Since there are several sections in the Wellman/Allied system where multiple trains may occur, it was decided to investigate this assumption in some detail.

Potential savings exist only in engineering costs and possibly field supervision, if the multiple units are constructed concurrently. Estimates were made indicating potential cost reductions are quite small. Multiple units of 2-6 trains show a possible reduction in investment of 1-3% compared with the basic assumption of multiple train costs.

6.7 Nomenclature

GP	Total flue gas to control plant	MACFM
GT	Total flue gas to each absorber train (maximum value of GT = 550)	MACFM
NA	Number of absorber trains	
SF	Total sulfur flow in flue gas to control plant	M lbs/hr.
S7	Total sulfur flow in flue gas to control unit per train of sulfur-related equipment in absorber and SO_2 regeneration areas (maximum value of S7 = 7).	M lbs/hr.
S28	Total sulfur flow in flue gas to control unit per equipment train in the purge/make-up area (maximum value of S28 = 28)	M lbs/hr.
N7	Number of trains of sulfur-related equipment in the absorber and SO_2 regeneration areas.	
N28	Number of equipment trains in the purge/make-up area	
E	Major equipment cost (direct material and subcontracts)	\$M
M	Field Materials Costs	\$M
L	Field Labor Costs	
A,S,P,	Letters following E,M,L A refers to absorber area S refers to SO_2 regeneration area P refers to purge/make-up area R refers to SO_2 reduction area No letter following refers to total for all areas	
IF	Particulate index (IF = 1 if particulates are present in flue gas. IF = 0 if particulates are absent)	
RB	Retrofit difficulty factor of each boiler	
RP	Retrofit difficulty factor of gas-related equipment in the absorber area which is not in parallel trains, i.e., the fuel oil system; assumed to be equal to the highest RB	

BARC	Bare cost of the control unit	$M
TPI	Total Plant Investment	$M
TCR	Total Capital Required	$M
CONTIN	Contingency	$M
AS	Annual cost of sodium carbonate	$M/Yr
AN	Annual cost of natural gas	$M/YR
AFA	Annual cost of filter aid	$M/YR
AE	Annual cost of electric power	$M/yr
AH	Annual cost of steam	$M/Yr
ACW	Annual cost of cooling water	$M/Yr
AW	Annual cost of process water	$M/Yr
AF	Annual cost of fuel oil	$M/Yr
ASC	Annual sulfur credit	$M/Yr
APS	Annual purge solids credit or debit	$M/Yr
CS	Purchase price of sodium carbonate	$/ton
CN	Purchase price of natural gas	$/MSCF
CFA	Purchase price of filter aid	$/ton
CE	Purchase (or transfer) price of electricity	mills/KWH
CH	Purchase (or transfer) price of steam	$/M lbs.
CCW	Cost of cooling water	$/M gal.
CW	Cost of process water	$/M gal.
CF	Purchase price of fuel oil	$/MM Btu
VSC	Unit value of sulfur (negative if credit)	$/long ton
VPS	Unit value of purge solids (negative if credit)	$/ton
TO	Total number of operators	
CO	Unit cost of operating labor	$/hr
LF	Load factor of the power plant	
AOC	Annual net operating cost	$M/Yr
TAC	Total annual production cost	$M/Yr
F	Location Factor	

TABLE 6.1

UNIT COSTS USED IN ILLUSTRATIVE EXAMPLES - WELLMAN/ALLIED STACK GAS SCRUBBING MODEL

Purchased Price of Sodium Carbonate ($/Ton)	40.00
Purchased Price of Filter-Aid ($/Ton)	50.00
Purchased Price of Natural Gas ($/MSCF)	0.50
Purchased Price of Electricity (Mils/KWHR)	8.00
Purchased Price of Steam ($/MLB)	0.50
Purchased Price of Cooling Water ($/MGal)	0.02
Purchased Price of Process Water ($/MGal)	0.20
Purchased Price of Fuel Oil ($/MMBtu)	0.80
Sulfur Credit ($/LT)	5.00
Unit Cost of Solid Disposal ($/Ton)	1.00
Average Hourly Wages Per Gulf Coast ($/Hr)	7.00
Interest on Capital During Construction (%)	12.00

TABLE 6.2

WELLMAN/ALLIED PROCESS AND COST MODEL

SUMMARY OF EQUATIONS

Capital Cost Model

$$EA = \sum_{n=1}^{NA} RB\left[726\ (GT/550)^{0.5} + 639\ (GT/550)^{0.9}\right]_n + 119\ RP\ (GP/3300)^{0.5}$$

$$+ \left[133\ (S7/7)^{0.5} + 127\ IF\ (S7/7)^{0.6}\right] N7 \qquad \$M$$

$$ES = \left[209\ (S7/7)^{0.5} + 618\ (S7/7)^{0.6} + 157\ (S7/7)^{0.9}\right] N7 \qquad \$M$$

$$EP = \left[525\ (S28/28)^{0.5} + 380\ (S28/28)^{0.6} + 86\ (S28/28)^{0.7}\right.$$

$$\left. + 306\ (S28/28)^{0.8} + 519\ (S28/28)^{0.9}\right] N28 \qquad \$M$$

$$ER = 998\ (SF/28)^{0.5} + 287\ (SF/28)^{0.6} + 683\ (SF/28)^{0.9} \qquad \$M$$

$$M = 0.429\ EA + 0.742\ ES + 0.827\ EP + 0.772\ ER \qquad \$M$$

$$L = 0.224\ EA + 0.310\ ES + 0.433\ EP + 0.623\ ER \qquad \$M$$

$$BARC = 1.15\ (E+M) + 1.43\ L \cdot F \qquad \$M$$

$$TPI = 1.12\ (1.0 + CONTIN)\ BARC \qquad \$M$$

$$TCR = 1.15\ TPI + 0.8\ TO \cdot CO\ (1+F) + 0.4\ ANR \qquad \$M$$

Operating Cost Model

$$AS = 28.2\ CS \cdot LF\ (SF/28) \qquad \$M/yr.$$

$$AN = 1460\ CN \cdot LF\ (SF/28) \qquad \$M/yr.$$

$$AFA = 1.24\ CFA \cdot LF \cdot IF\ (GP/3300) \qquad \$M/yr.$$

$$AE = \left[154\ (GP/3300) + 79\ (SF/28)\right] CE \cdot LF \qquad \$M/yr.$$

$$AH = 5430\ CH \cdot LF\ (SF/28) \qquad \$M/yr.$$

$$ACW = \left[856\ (GP/3300) + 19{,}900\ (SF/28)\right] CCW \cdot LF \qquad \$M/yr.$$

$$AW = 64\ (SF/28)\ CW \cdot LF \qquad \$M/yr.$$

TABLE 6.2 (Cont'd)

AF = 1,800 (GP/3300) CF·LF $M/yr.

ASC = 95.4 (SF/28) VSC·LF $M/yr.

APS = 37.3 (SF/28) VPS·LF $M/yr.

ANR = AS + AAO + AN+ AFA + AE + AH + ACW
+ AW + AF + ASC + APS $M/yr.

AOC = 0.078 TPI + 2·TO·CO (1+F) + ANR $M/yr.

TAC = 0.237 TPI + 2.1·TO·CO (1+F) + 1.04 ANR $M/yr.

FIGURE 6.1

WELLMAN/ALLIED PROCESS FLOWSHEET

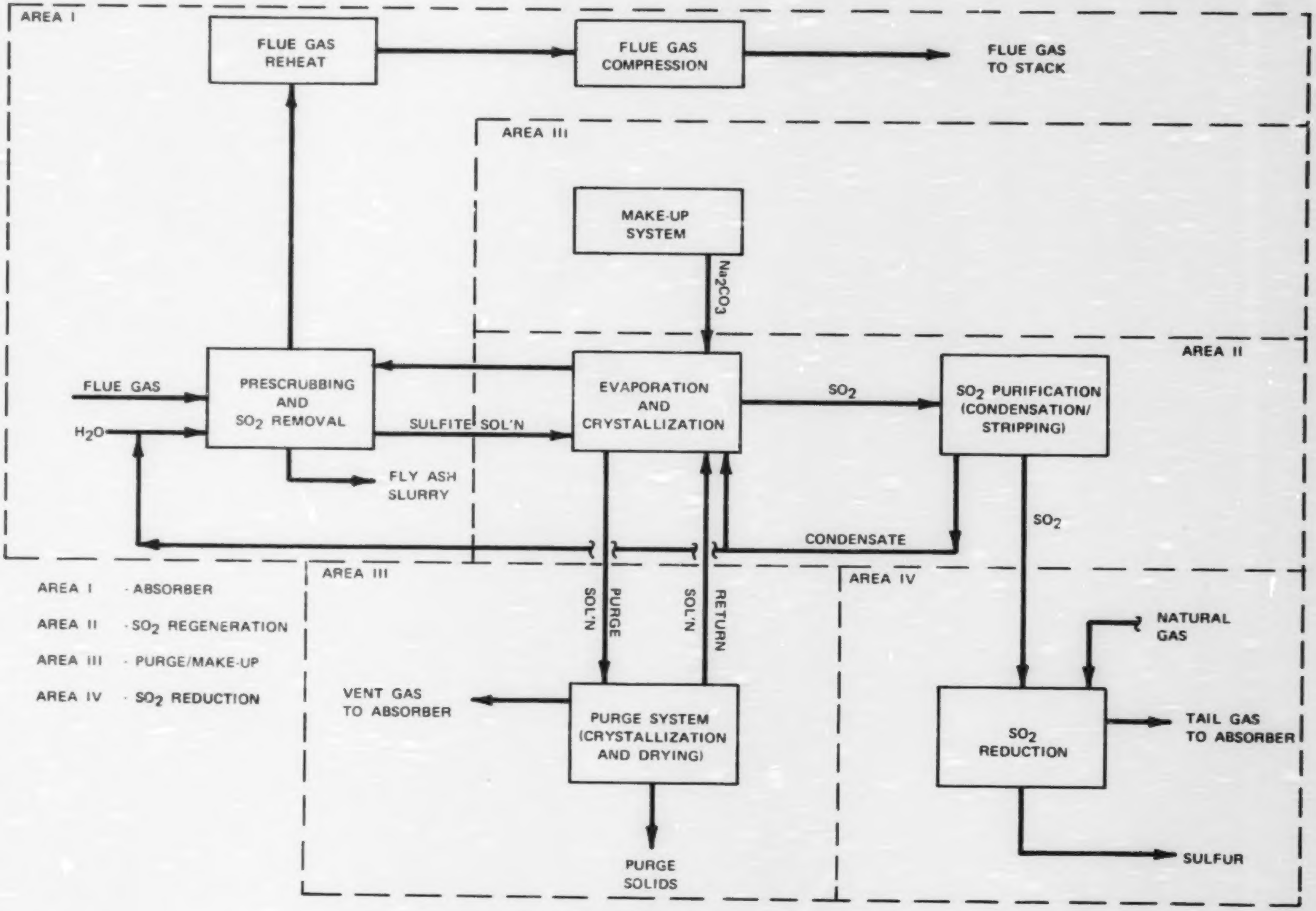
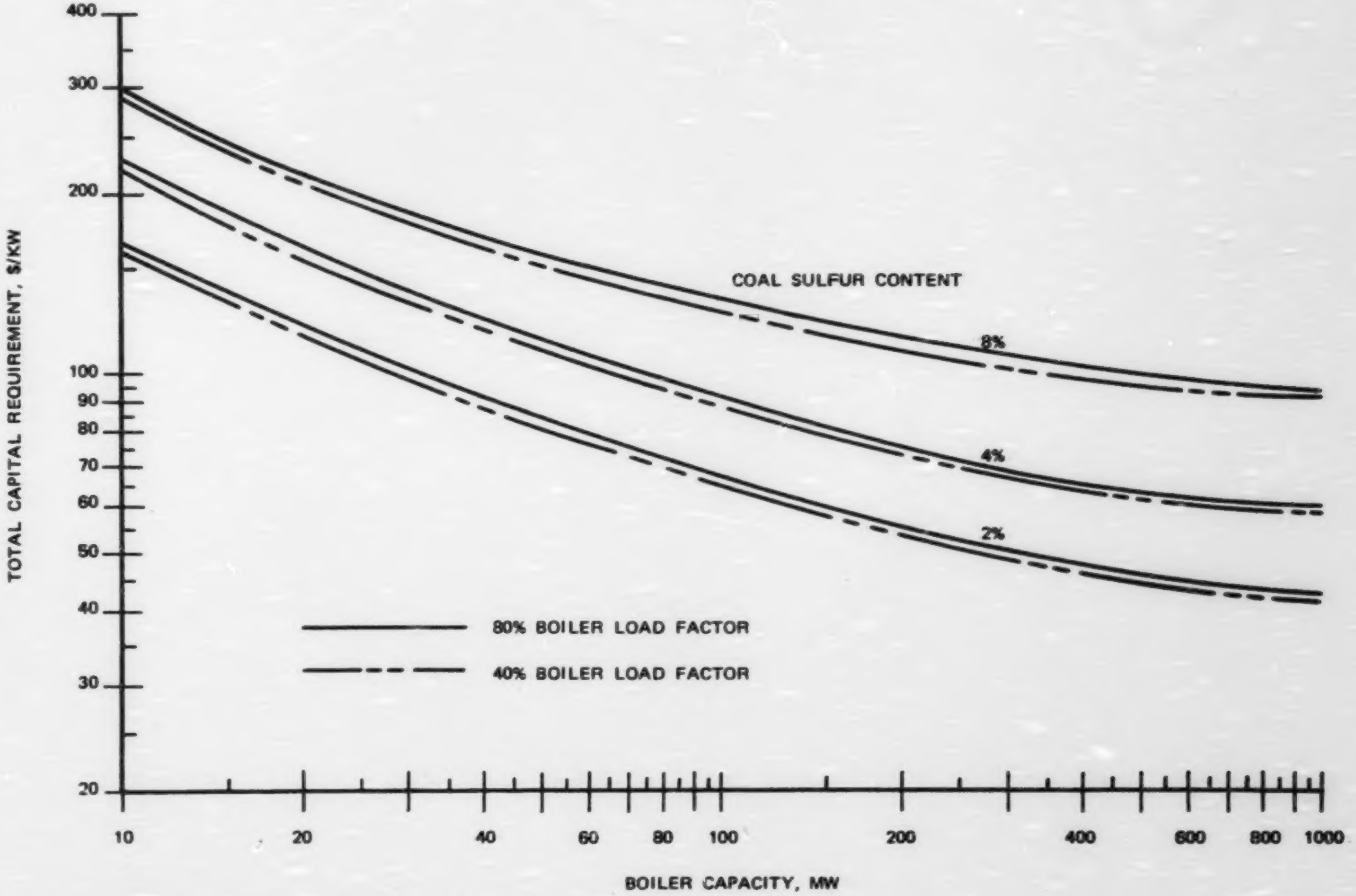

AREA I
FLUE GAS REHEAT
FLUE GAS COMPRESSION
FLUE GAS TO STACK
AREA III
MAKE-UP SYSTEM
Na_2CO_3
AREA II
FLUE GAS
H_2O
PRESCRUBBING AND SO_2 REMOVAL
SULFITE SOL'N
EVAPORATION AND CRYSTALLIZATION
SO_2
SO_2 PURIFICATION (CONDENSATION/ STRIPPING)
FLY ASH SLURRY
CONDENSATE
SO_2
AREA III
PURGE SOL'N
RETURN SOL'N
AREA IV
NATURAL GAS
VENT GAS TO ABSORBER
PURGE SYSTEM (CRYSTALLIZATION AND DRYING)
SO_2 REDUCTION
TAIL GAS TO ABSORBER
PURGE SOLIDS
SULFUR
AREA I - ABSORBER
AREA II - SO_2 REGENERATION
AREA III - PURGE/MAKE-UP
AREA IV - SO_2 REDUCTION

FIGURE 6.2

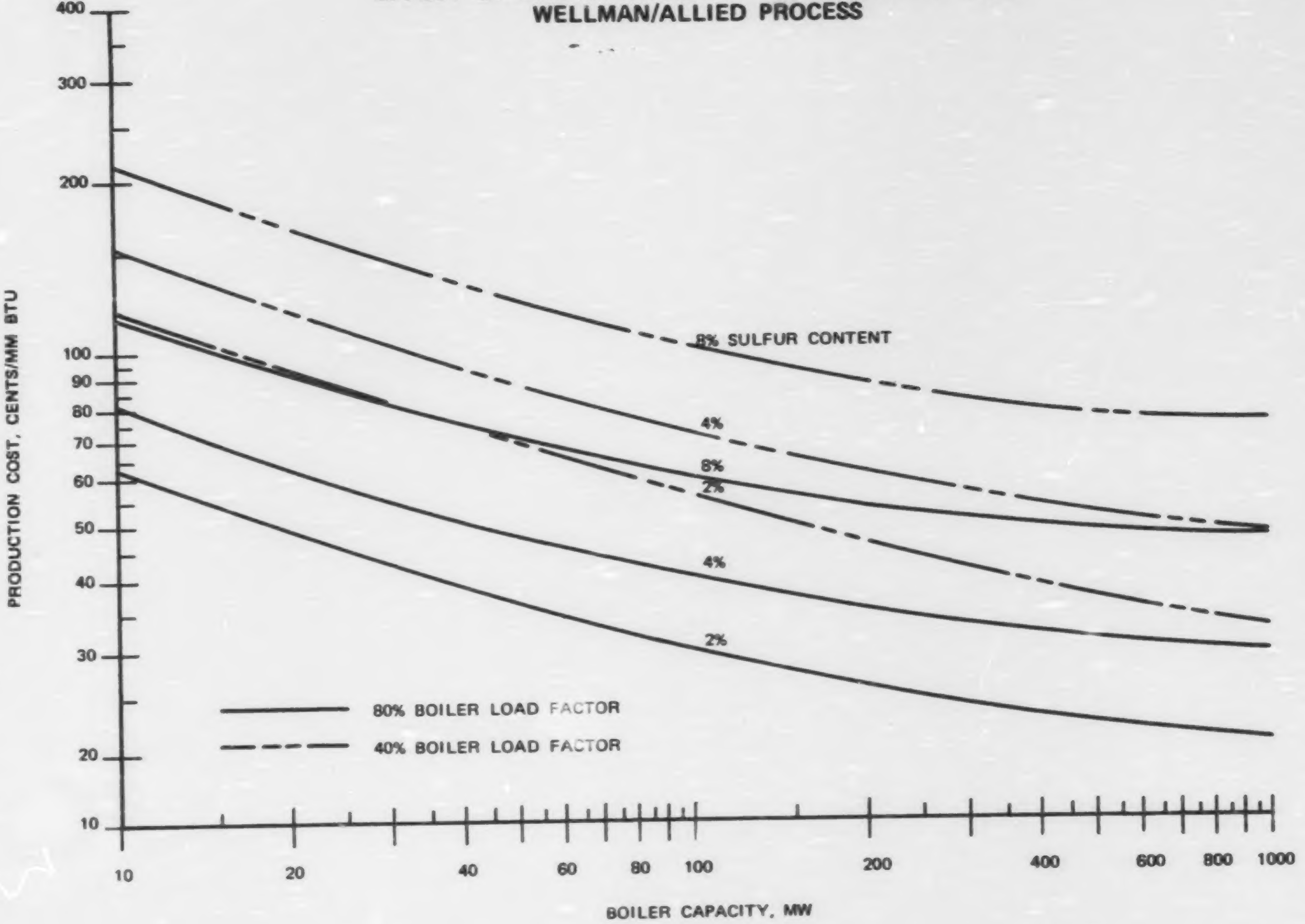
EFFECT OF BOILER CAPACITY ON TOTAL CAPITAL REQUIREMENT
WELLMAN/ALLIED PROCESS



BASIS OF CALCULATION: NO RETROFIT, NO CONTINGENCY, U.S. GULF COAST LOCATION, END OF 1973 FIGURE, BOILER HEAT RATE 9,500 BTU/KWH, HEATING VALUE OF COAL (HHV) 11,000 BTU/LB.

FIGURE 6.3

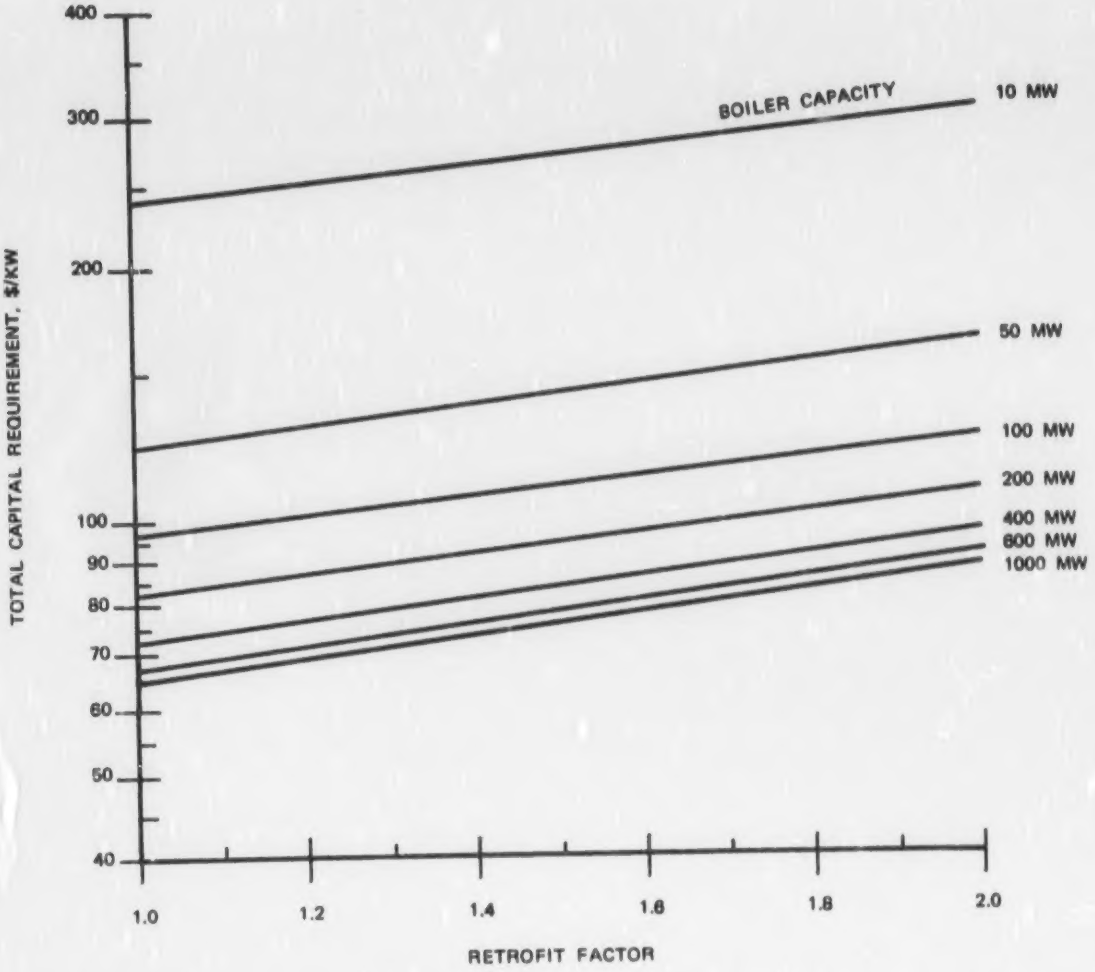
EFFECT OF BOILER CAPACITY ON PRODUCTION COST WELLMAN/ALLIED PROCESS



(SEE FIGURE 6.2 FOR BASIS OF CALCULATION)

FIGURE 6.4

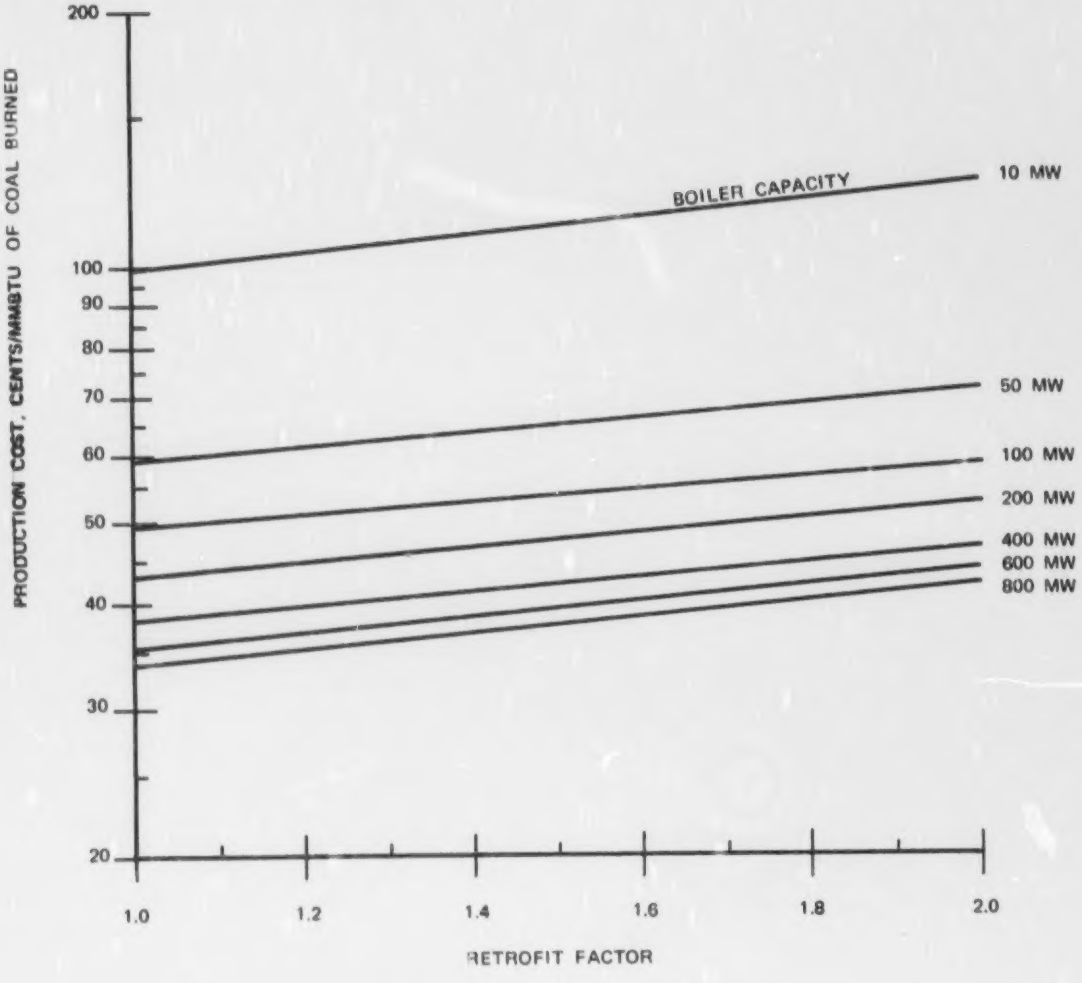
EFFECT OF BOILER RETROFIT DIFFICULTY ON TOTAL CAPITAL REQUIREMENT WELLMAN/ALLIED PROCESS



BASIS OF CALCULATION: 4% SULFUR COAL, HEATING VALUE OF 11,000 BTU/LB, INDIVIDUAL BOILER WITH HEAT RATE OF 9,500 BTU/KWH AND LOAD FACTOR OF 0.7, CONTINGENCY 10%, U.S. GULF COAST LOCATION, END OF 1973 FIGURE.

FIGURE 6.5

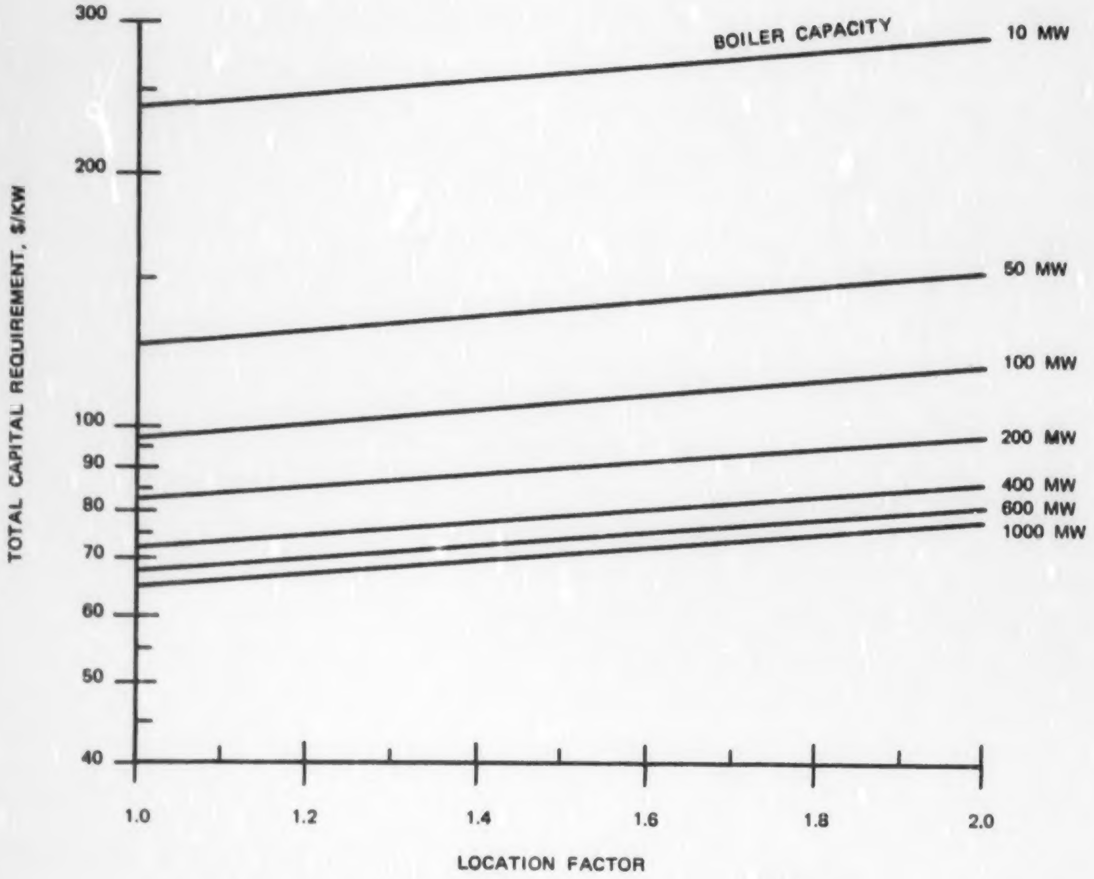
EFFECT OF BOILER RETROFIT DIFFICULTY ON PRODUCTION COST WELLMAN/ALLIED PROCESS



(SEE FIGURE 6.4 FOR BASIS OF CALCULATION.)

FIGURE 6.6

EFFECT OF LOCATION FACTOR ON TOTAL CAPITAL REQUIREMENT WELLMAN/ALLIED PROCESS



BASIS OF CALCULATION: 4% SULFUR COAL, HEATING VALUE OF 11,000 BTU/LB, INDIVIDUAL BOILER WITH HEAT RATE OF 9,500 BTU/KWH AND LOAD FACTOR OF 0.7, CONTINGENCY OF 10%, END OF 1973 FIGURE.

7. APPLICATION OF STACK GAS SCRUBBING MODELS

7.1 Stack Gas Scrubbing Applied to Existing Utilities

One of the main concerns in this study was to investigate the cost of retrofitting stack gas scrubbing units, which have been described in the previous sections, to existing utility boilers. This exercise was carried out for the existing utilities in 1971 using the data from the statistics discussed in section 3. The cost of retrofitting stack gas scrubbing units was analyzed and averaged for the different plant sizes on a national basis.

A program was written first to analyze the existing data. These data include plant size, the types and amounts of fuel burned in the plant, the heating value of the respective fuels, the number of boilers, boiler sizes, etc. If part of these data was missing, the plant data were upgraded by the program using the average statistics for that size plant. Table 7.1 presents part of these statistics. The number of boilers, the percent of the plant capacity attributable to the largest boiler size, the percent attributable to the second largest boiler size etc., for various size ranges of utility plants are given. The average number of boilers per plant is fairly constant and is equal to four for all cases studied except plants below 50 megawatts size. The figures presented are the averages for U. S. utilities. There may be some exceptional cases where the fifth and sixth boilers are still significant.

Figure 7.1 represents a highly realistic situation in which the largest boiler is base loaded and the load factor decreases with size. The smaller boilers are used for peaking only. The load factor is defined as follows:

$$\text{LOAD FACTOR} = \frac{\text{Total Yearly Generation}}{\text{Maximum Capacity} \times 8760 \text{ hours/yr}}$$

If the load factor for the largest boiler is unity, the load factor for the second largest boiler is about 0.75. The load factor for the third boiler is 0.55 and the remaining boiler or boilers are below 0.15. The procedure for calculating individual boiler load factor has been set up to give the exact load factor for the plant.

Figure 7.2 shows the average heat rate of a boiler versus boiler size. The smaller boilers are generally the older ones and consequently less efficient than the newer, larger boilers. In this exercise an average heat rate of 13,000 Btu/kwh, representing an overall cycle efficiency of 26%, has been taken as the upper limit, and an average heat rate of 9,000 Btu/kwh, representing an overall cycle efficiency of 38%, as the lower limit. Any value outside this range is viewed as an error in the data collected and has been adjusted before the actual calculations. Using the boiler load factor and the boiler heat rate, the fuel demand for the boilers can be calculated for the utilities. Fuel is then allocated in the order of coal, oil and gas starting with the largest boiler and working toward the smallest boiler. The next step is to investigate the overall plant SO_2 emission. If the overall plant SO_2 emission is above the specified level of 1.2 lb/MMBtu, a stack gas scrubbing unit (either the Wet Limestone or the Wellman/Allied) is fitted first to the largest sulfur emitting boiler, then the second largest, then the third largest etc. until the overall plant SO_2 emission is below the emission level specified. The cost for installing stack gas scrubbing units described in this manner should be fairly realistic and represents the minimum cost in terms of $/KW of plant capacity. Figures 7.3 to 7.14 present graphically the results from this exercise.

Figures 7.3 and 7.9 show the total capital required for installing Wet Limestone and Wellman/Allied stack gas scrubbing units in existing utilities of various sizes. Cost differences between the two processes are quite small and well within the order of accuracy of the models.

Figures 7.4 and 7.10 present the total capital required, expressed as \$/KW of plant capacity, for the two processes. The cost in terms of \$/KW of plant capacity increases gradually with decreasing plant size from 2,000 megawatts to 100 megawatts but rises sharply below 100 megawatts plant size.

Figures 7.5 and 7.11 show the estimated incremental cost of electricity delivered in mils/kwh for installing stack gas scrubbing in the existing utilities. For the two processes studied, the incremental cost varies from about 5 mils/kwh to 1.5 mils/kwh for plant sizes of 100 megawatts to 1500 megawatts. Below 100 megawatts size, the incremental cost rises sharply with decrease in size. It would be practical to require utility plants to be fitted with Wet Limestone or Wellman/Allied process stack gas scrubbing units if the incremental cost of electricity delivered is less than 4 mils/kwh. Above this value, other alternatives, such as burning clean fuel, should be investigated if the control on SO_2 emissions is to be imposed.

Figures 7.6 and 7.12 show the total cummulative demand for clean fuel versus the incremental cost which could be paid for the fuel as an alternate to stack gas scrubbing, and the range of plant sizes in which the clean fuel would be burned. For both processes, the conclusions are about the same. If clean fuel is available at an incremental cost below \$0.30/MMBtu, there is a potential market of 6×10^9 MMBtu/year. However, if the clean fuel is available at an incremental cost above \$2.50/MMBtu, the potential market decreases to 1.50×10^8 MMBtu/year, corresponding to a reduction of 97.5%.

Figures 7.7, 7.8, 7.13, 7.14 show the relationship between the average cost of stack gas scrubbing for the existing utilities (starting with the largest plants) and the cumulative percent of total U.S. capacity under control. For the Wet Limestone process, the average cost for controlling 10% of the U.S. capacity to meet the emission standard of 1.2 lb SO_2 per MMBtu of fuels burned is \$40/KW of plant capacity and 1.5 mil/Kwh of electricity delivered. To control 100% of the U.S. capacity, the average cost increases to \$64/KW of plant capacity and 2.8 mils/Kwh of electricity delivered. For the Wellman/Allied process, the corresponding figures are \$42/KW and 1.7 mils/Kwh for controlling 10% of the U.S. capacity; \$68/KW and 3.0 mils/Kwh for controlling 100% of capacity to meet the emission standard.

It must be stressed that the figures presented here are not the cost for controlling the total plant capacity but rather controlling enough boilers to meet the specified SO_2 emission level. The location factor, which is described in the General Cost Model (section 4), is incorporated into the calculations and the figures are the average costs for the U.S. utilities on a national basis. The overall conclusion is that it is economically preferable to install stack gas scrubbing units on larger size utility plants while for small size utility plants (below 100 megawatts) the better alternative is to burn low sulfur fuel.

7.2 Stack Gas Scrubbing Applied to Industrial Boilers

An investigation was made to determine the costs of fitting stack gas scrubbing processes to coal- and oil-fired industrial boilers in the United States. As an initial phase, the process and cost models for the Wet Limestone and Wellman/Allied processes were reviewed to determine their applicability to industrial boilers. Of particular interest were the smaller boilers, since these represent a large extrapolation of the models from the type of application for which they were initially developed, viz., large utility boilers.

As a result of this review, a number of changes were incorporated into the Wet Limestone and Wellman/Allied models. These changes are briefly discussed below.

7.2.1 Wet-Limestone Process

A review of the Wet Limestone model prompted the following changes:

1) Replacement of the sludge pond with a thickener and temporary sludge disposal pit.

2) Elimination of onsite limestone grinding at small limestone design rates (low sulfur flows).

3) Reduction of some of the scrubbing equipment costs for small boiler sizes.

It was felt that, for industrial boiler applications, a thickener circuit for sludge handling would be more universally applicable than a large sludge pond. The sludge would be periodically hauled offsite and disposal treated as an operating cost. Grinding of limestone becomes increasingly expensive as the limestone design rate decreases, and it was assumed that at sulfur flows of less than 2000 lbs/hr, grinding would be eliminated in favor of purchasing pulverized stone. Costs for some of the scrubbing equipment were found to be high for boiler sizes less than about 400 MM BTU/hr and were reduced accordingly.

Table 7.2 summarizes the changes made to the equipment cost portion of the model. The equation for chemical process equipment costs, EC, now includes the factor FC which reduces the cost of some of the scrubbing equipment for small boilers. This factor varies with boiler capacity as shown, ranging in value from 1.0 to 2.25. The term ISF is an index used to delete the grinding equipment costs when the sulfur rate falls below 2000 lbs/hr. P now represents the cost of the small temporary storage pit for limestone sludge.

The raw materials and utilities cost equations are presented in Table 7.3. The first equation represents the annual cost of sludge disposal, ASL, in terms of the unit cost, CSL, in \$/ton. The equation for the annual cost of limestone, AL, remains unchanged in form. However, when pulverized limestone is purchased (i.e., when grinding is eliminated), the numerical value of the unit cost of limestone, CL, would be increased appropriately. The last equation in the table gives electric power costs and now includes a term which reduces the process power consumption when limestone grinding is eliminated.

In Figures 7.15-7.18 typical costs which were calculated from the cost model have been plotted to illustrate the effects of different variables on the Wet-Limestone stack gas scrubbing costs. The figures have been separated into 10-100 MM Btu/hr and 100-1000 MM Btu/hr size ranges because of difficulty in scaling.

Figures 7.15 and 7.16 illustrate the effect of boiler capacity on the total capital required (TCR) with load factor and sulfur content of coal shown as parameters. For small boilers (10-100 MM Btu/hr), the load factor and percent sulfur have insignificant effects on capital required. The effect becomes noticeable when the boiler size becomes larger. For a boiler size of 1000 MM Btu/hr, doubling the sulfur content in coal (from 2% to 4% or from 4% to 8%) increases the capital required by approximately 10%. The effect of load factor remains minor.

Figure 7.17 and 7.18 illustrate the pronounced effects of sulfur content and load factor on the operating cost (TAC). Generally speaking, doubling the sulfur content in coal increases the operating cost by 10% for the small boilers (10-50 MM Btu/hr) and the percent gradually increases as boiler capacity increases, to as much as 20% for 1000 MM Btu/hr. The operating cost can be increased by as much as 100% for smaller boilers (10 MM Btu/hr) when the load factor is reduced to half (0.8 to 0.4) and as much as 60% for larger boilers (1000 MM Btu/hr).

7.2.2 Wellman/Allied

A review of the Wellman/Allied model suggested that two changes could be made for application to industrial boilers. First, it was found that the predicted costs of some of the scrubbing equipment were high for small boiler sizes, as in the Wet Limestone model. It was also found that the same adjustment factor used in the Wet Limestone model could be used in the Wellman/Allied model.

The second change would affect the SO_2 regeneration and sulfur recovery areas. Most industrial boilers operate at a fairly low load factor, indicating a significant variation in their operating rate throughout the year. Since their operation is tied exclusively to a particular plant or plant site, this variation in operating rate might be more reliably predicted than for a utility plant which is tied into a grid system. Consequently, it might be possible for industrial boiler applications to size the regeneration and recovery areas for somewhat less than peak sulfur load by providing adequate surge capacity between the absorber and regeneration plant. For purposes of illustrating the effect of this type of design on costs, a design point 25% above the average sulfur flow has been assumed adequate for the regeneration plant. Surge capacity has been increased accordingly.

The resultant equipment cost equations are summarized in Table 7.6. The first equation now includes the term, FC, to reduce some of the scrubbing equipment costs for small boiler applications. The primed variables (S7', S28', SF') in the equations reduce the size and cost of the regeneration and sulfur recovery plant and are related in the same manner as the unprimed variables (see Section 6).

In Figures 7.19-7.22 typical costs which were calculated from the cost model have been plotted to illustrate the effects of different variables on the Wellman/Allied stack gas scrubbing process. The figures have been separated into 10-100 MM Btu/hr and 100-2500 MM Btu/hr size ranges.

Figures 7.19 and Figures 7.20 illustrate the effect of boiler capacity on the total capital required (TCR), with percent sulfur and load factor as parameters. To some extent, the load factor and percent sulfur have a larger effect on the capital required in the Wellman/Allied system than in the Wet Limestone process. Generally speaking, doubling the percent sulfur in coal or decreasing the load factor to half increases the total capital required by 20% to 40%; the effect is more pronounced for smaller boilers (10-100 MM Btu/hr) than for large boilers (100-2500 MM Btu/hr).

Figures 7.21 and 7.22 illustrate the effect of load factor and percent sulfur in coal on the operating costs (TAC). It can be seen from these figures that doubling the sulfur percent in coal increases the operating cost by about 30% for large boilers (100-2500 MM Btu/hr) and by as much as 50% for small boilers. The effect of load factor is more pronounced than percent sulfur with the increase in operating cost, for decreasing the load factor to half, ranging from 50% to 70%. Again the effect is more pronounced for small boilers (10-100 MM Btu/hr).

7.2.3 Applicability to Small Industrial Boilers

Coal- and oil-fired industrial boilers in the United States number more than 5,000. Based on the statistical analysis of the available boiler population, these range in capacity up to 4400 MM Btu/hr. The small boilers, i.e., those with a capacity of 100 MM Btu/hr or less, represent almost three-fourths of the total population. However, these boilers emit less than one-fourth of the total sulfur emissions from all coal- and oil-fired industrial boilers.

Figures 7.19-7.22, which show typical costs of stack gas scrubbing, indicate that the costs incurred by small size industrial boilers ($\leq$ 100 MM Btu/hr) are very high. Depending on boiler size, clean fuel at incremental prices of roughly \$1-3/MM Btu (or less) would be preferable to scrubbing as a control method.

However, considering the limited emissions from these small boilers, it would be difficult to justify either type of control unless there were particularly good local reasons.

The preceding discussion is based on single boiler installations. There were no data available to permit estimation of costs on a plant basis (i.e., considering the total number of boilers per plant). This type of analysis should be done to obtain more meaningful costs.

TABLE 7.1 BOILER SIZE DISTRIBUTION FOR STANDARD SIZE UTILITY PLANT

PLANT SIZE	AVERAGE NO. OF BOILERS	AVERAGE DISTRIBUTION-PLANT GENERATION CAPACITY, %			
		BOILER 1	BOILER 2	BOILER 3	BOILER 4
Megawatts					
0 to 10	1	100.0%	-	-	-
11 to 50	3	65.0	26.0	9.0	-
51 to 100	4	58.0	24.0	12.0	6.0
101 to 200	4	57.0	25.0	11.0	7.0
201 to 300	4	53.0	27.0	12.0	8.0
301 t0 400	4	52.0	27.0	13.0	8.0
401 to 500	4	53.0	26.0	14.0	7.0
501 to 600	4	58.0	23.0	12.0	7.0
601 to 700	4	56.0	27.0	11.0	6.0
701 to 800	4	49.0	34.0	12.0	5.0
801 to 900	4	58.0	25.0	9.0	7.0
901 to 1000	4	52.0	28.0	14.0	6.0
1001 to 1200	4	47.0	30.0	13.0	10.0
1201 to 1600	4	46.0	35.0	10.0	9.0
1401 to 1600	4	45.0	28.0	20.0	7.0
Over 1600	4	41.0	38.0	13.0	8.0

TABLE 7.2

WET LIMESTONE PROCESS

SUMMARY OF EQUIPMENT COST EQUATIONS FOR INDUSTRIAL BOILERS

$$EC = \sum_{n=1}^{NA} RB \left[\left(629 + \frac{412}{FC}\right) (GT/550)^{0.5} + \left(200 + \frac{208}{FC}\right) (GT/550)^{0.9} \right]_n$$

$$+ 238\ RB\ (GP/3300)^{0.5} + (201-7ISF)\ (SF/28)^{0.5} \qquad \$M$$

$$ES = (1680-1180\ ISF)\ (SF/28)^{0.9} + 120\ (SF/28)^{0.7} \qquad \$M$$

$$P = 40\ (SF/28)^{0.5} \qquad \$M$$

$FC = 2.25-0.003\ CAP \quad (FC \geq 1.0)$

where CAP = boiler capacity, MM BTU/hr input

$ISF = 1$ when $SF < 2$

$ISF = 0$ when $SF \geq 2$

TABLE 7.3

WET LIMESTONE PROCESS

SUMMARY OF OPERATING COST EQUATIONS FOR INDUSTRIAL BOILERS

RAW MATERIALS AND UTILITIES

$ASL = 2210 \cdot CSL \cdot LF \cdot (SF/28)$	\$M/year
$AL = 600 \cdot CL \cdot LF \cdot (SF/28)$	\$M/year
$AA = 0.43 \cdot CA \cdot (SF/28)$	\$M/year
$AF = 1800 \cdot CF \cdot LF \cdot (GP/3300)$	\$M/year
$AW = 230 \cdot CW \cdot LF \cdot [(GP/3300) + (SF/28)]$	\$M/year
$AE = CE \cdot LF \cdot [213(GP/3300) + (35-23.5\ ISF)\ (SF/28)]$	\$M/year

TABLE 7.4

WELLMAN/ALLIED PROCESS

SUMMARY OF EQUIPMENT COST EQUATIONS FOR INDUSTRIAL BOILERS

$$EA = \sum_{n=1}^{NA} RB\left[\left(407 + \frac{319}{FC}\right)(GT/550)^{0.5} + \left(200 + \frac{439}{FC}\right)(GT/550)^{0.9}\right]_n$$

$$+ 119\ RB\ (GP/3300)^{0.5}$$

$$+ \left[190\ (S7/7)^{0.5} + 50\ (S7'/7)^{0.5} + 127IF\ (S7'/7)^{0.6}\right]N7 \qquad \$M$$

$$ES = \left[209\ (S7'/7)^{0.5} + 618\ (S7'/7)^{0.6} + 157\ (S7'/7)^{0.9}\right]N7 \qquad \$M$$

$$EP = \left[525\ (S28'/28)^{0.5} + 380\ (S28'/28)^{0.6} + 86\ (S28'/28)^{0.7}\right.$$

$$\left. + 306\ (S28'/28)^{0.8} + 519\ (S28'/28)^{0.9}\right]N28 \qquad \$M$$

$$ER = 998\ (SF'/28)^{0.5} + 287\ (SF'/28)^{0.6} + 683\ (SF'/28)^{0.9} \qquad \$M$$

$FC = 2.25 - 0.003\ CAP\ (FC \geq 1.0)$
where CAP = boiler capacity, MM BTU/hr input

$SF' = SF$ if $LF > 0.8$
$SF' = 1.25 \cdot LF \cdot SF$ if $LF \leq 0.8$

FIGURE 7.1

AVERAGE DISTRIBUTION OF LOAD FACTORS FOR BOILERS IN A UTILITY PLANT

LOAD FACTOR OF BOILER N = RATIO X LOAD FACTOR OF BOILER 1

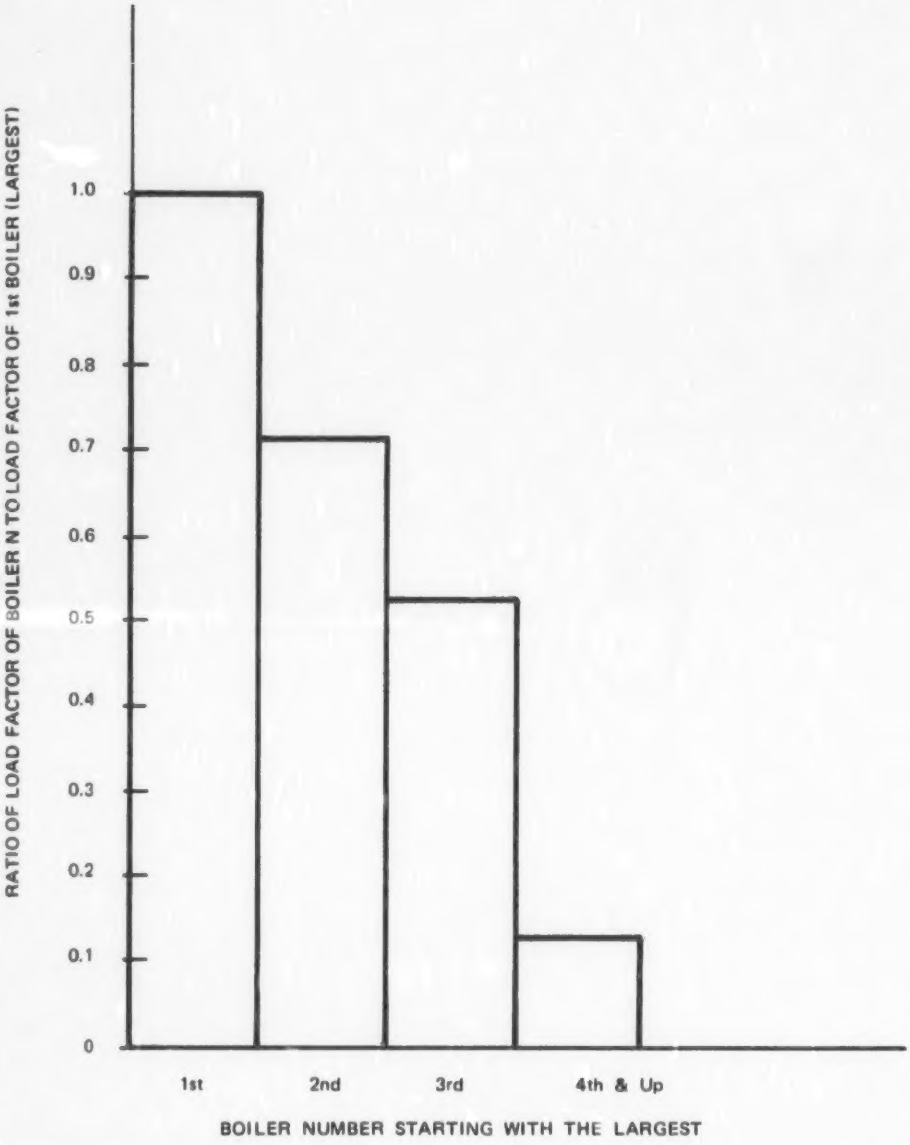


FIGURE 7.2

AVERAGE HEAT RATES FOR UTILITY BOILERS

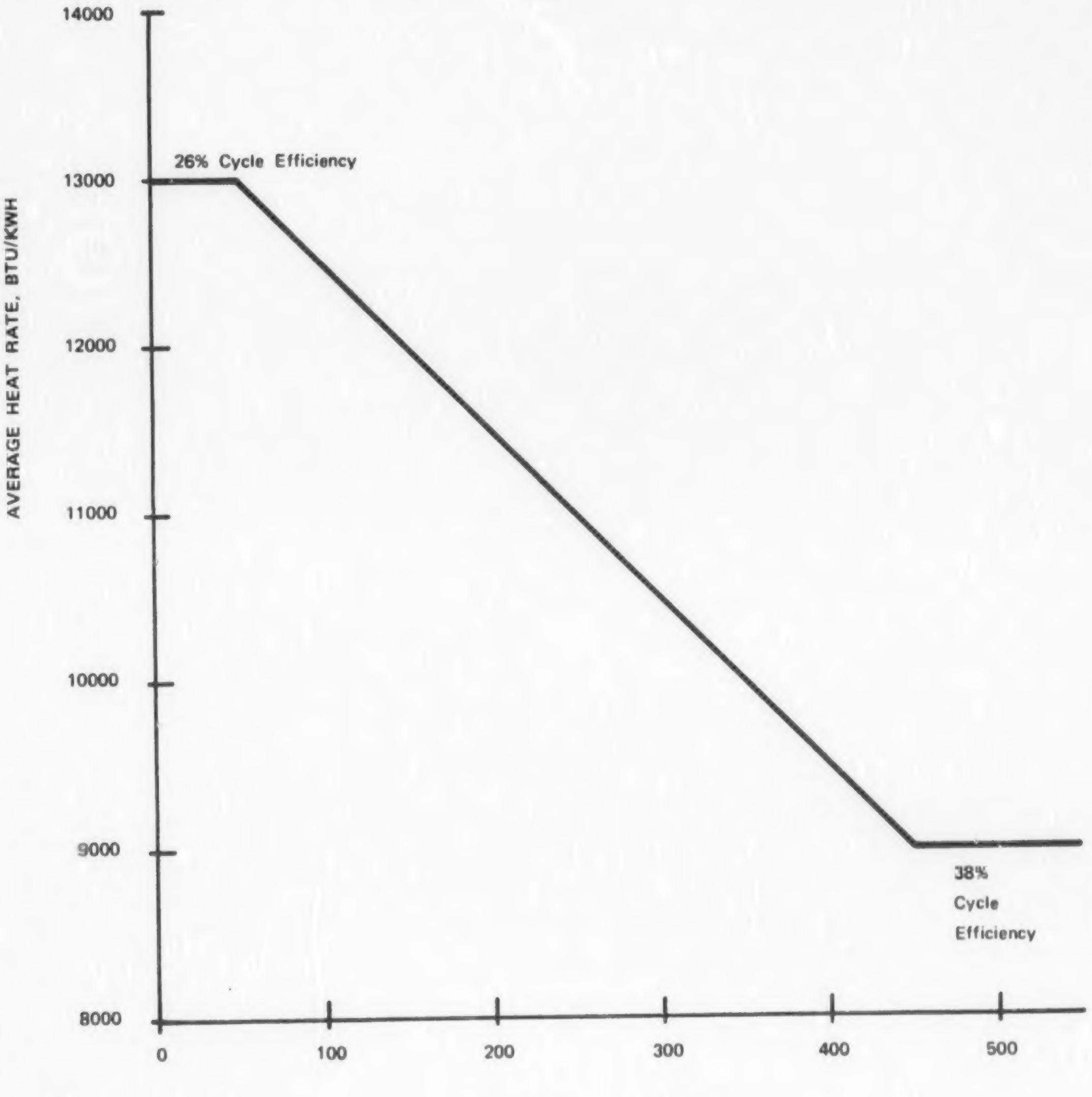


FIGURE 7.3

AVERAGE TOTAL CAPITAL REQUIREMENT FOR INSTALLING WET LIMESTONE SYSTEM IN EXISTING POWER PLANTS

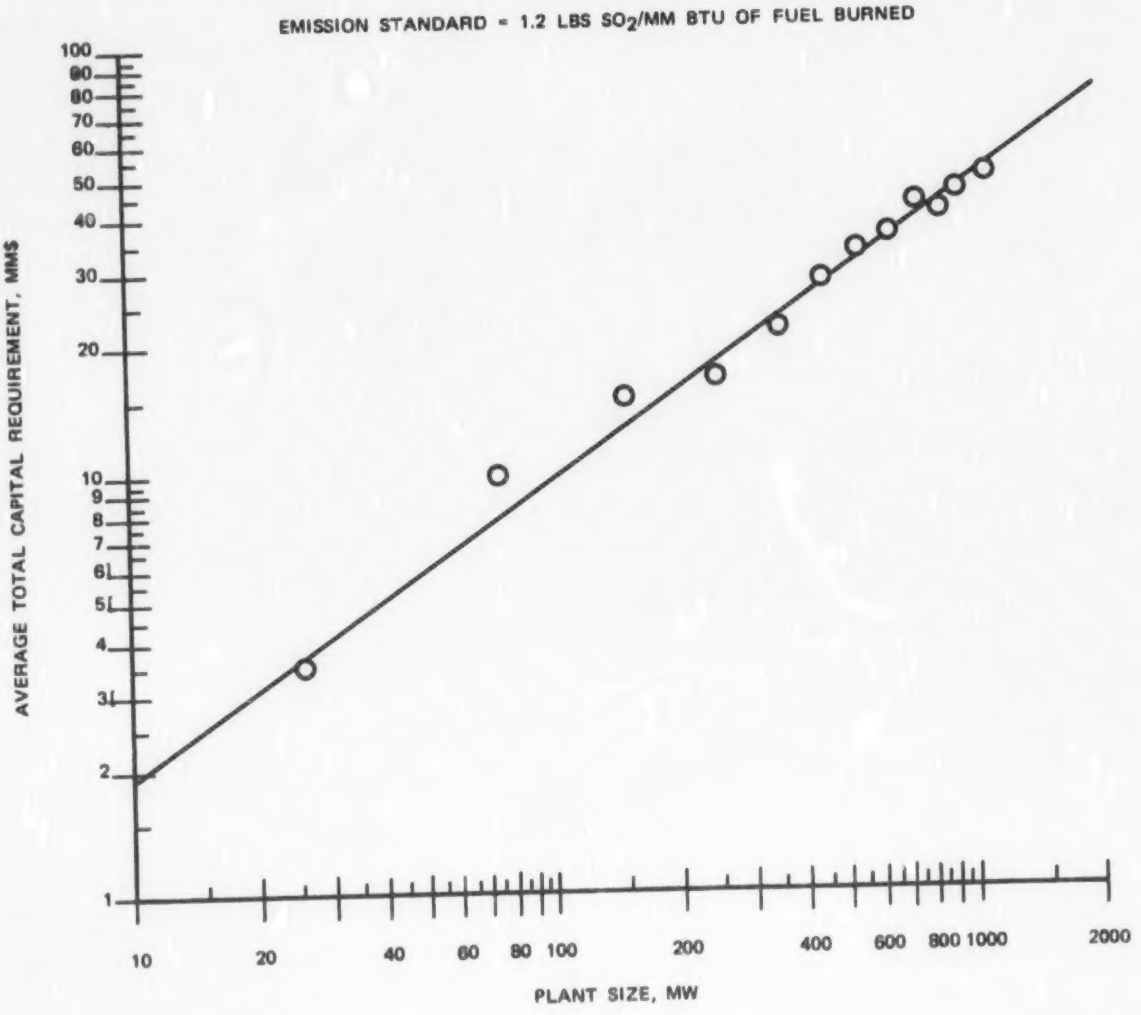


FIGURE 7.4

AVERAGE UNIT COST FOR INSTALLING WET LIMESTONE SYSTEM IN EXISTING POWER PLANTS

EMISSION STANDARD = 1.2 LBS SO_2/MM BTU OF FUEL BURNED

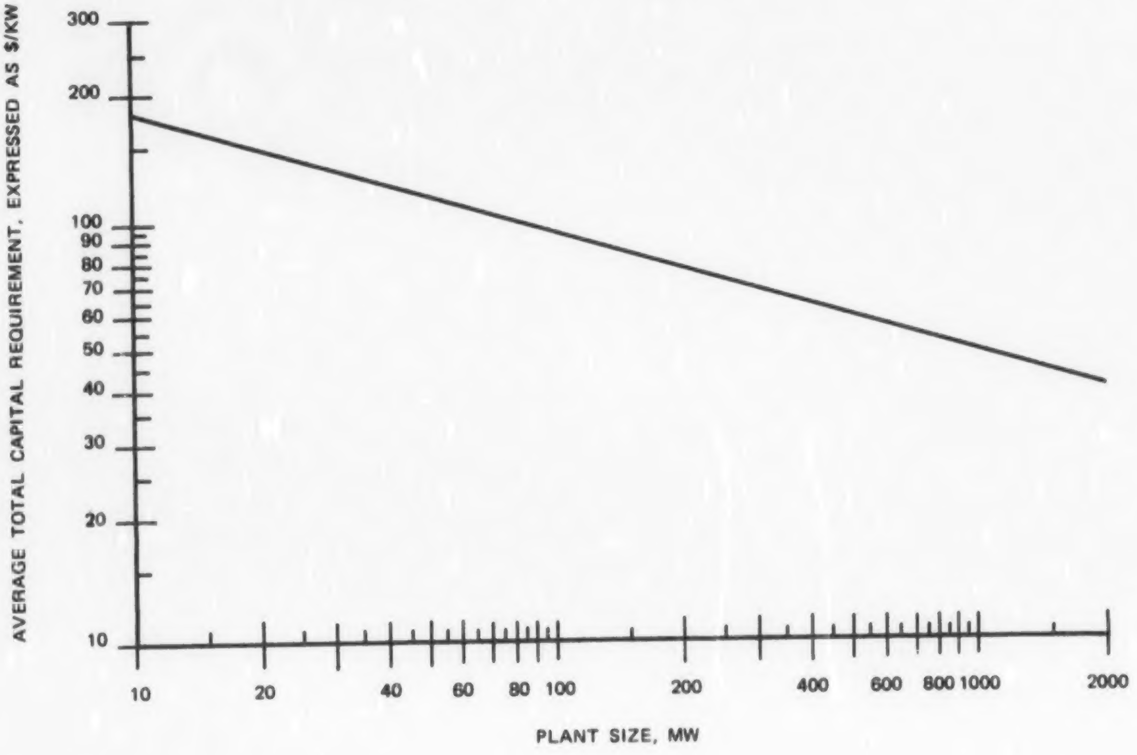


FIGURE 7.5

INCREMENTAL OPERATING COST FOR WET LIMESTONE SYSTEM IN EXISTING POWER PLANTS

EMISSION STANDARD = 1.2 LBS SO_2/MM BTU OF FUEL BURNED

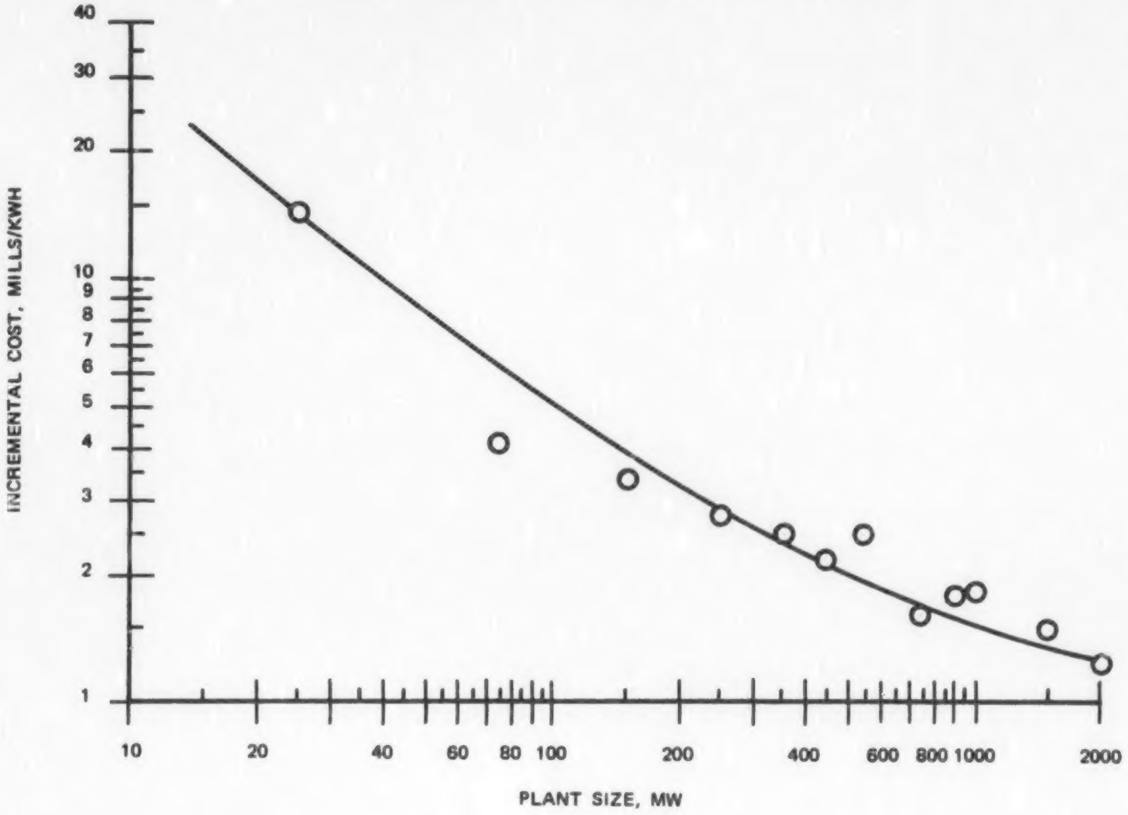


FIGURE 7.6

DEMAND FOR CLEAN FUEL AS ALTERNATIVE TO STACK GAS SCRUBBING WET LIMESTONE SYSTEM APPLIED TO EXISTING POWER PLANTS

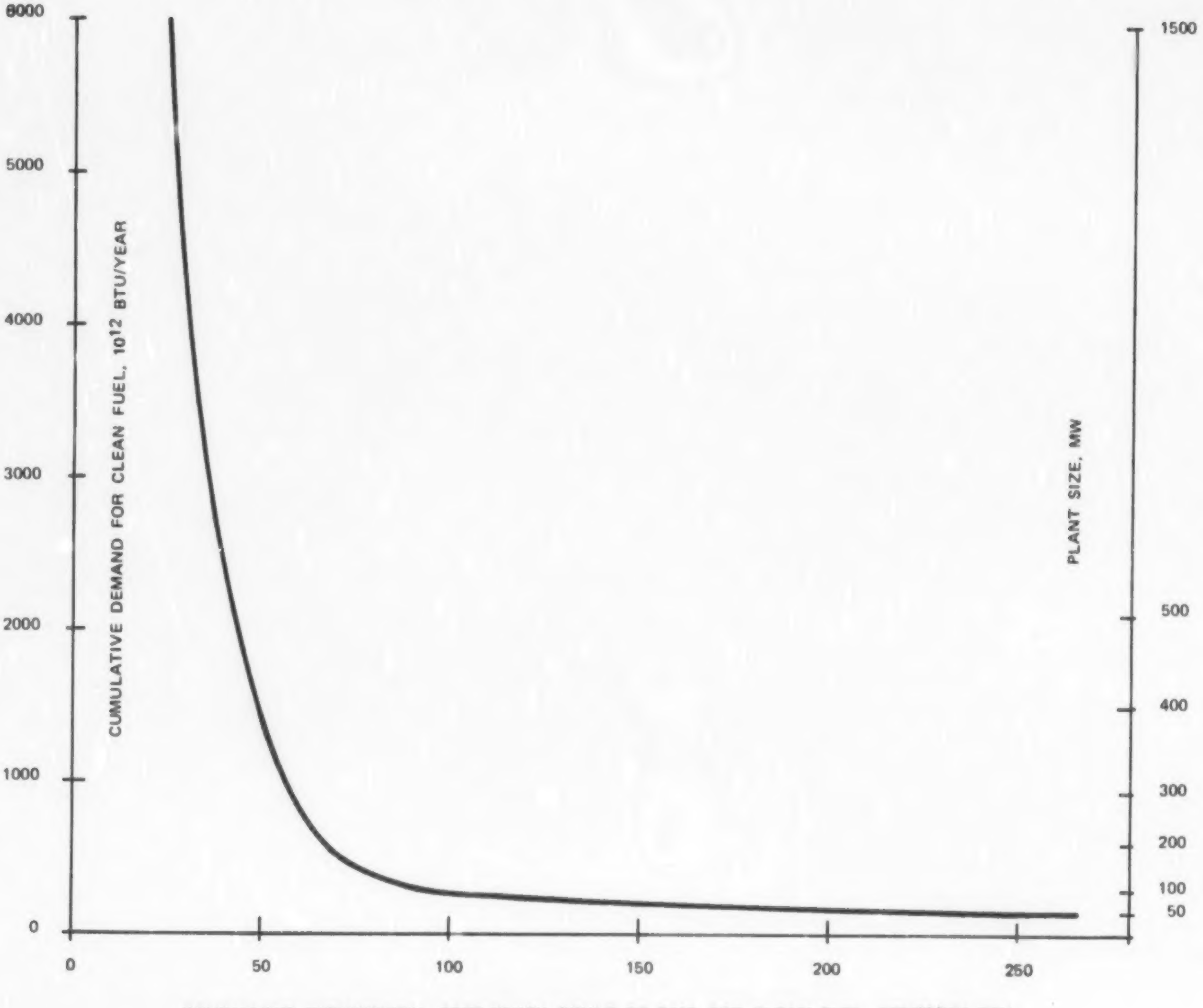


FIGURE 7.7

CUMULATIVE AVERAGE CAPITAL COST
WET LIMESTONE SYSTEM APPLIED TO POWER PLANTS

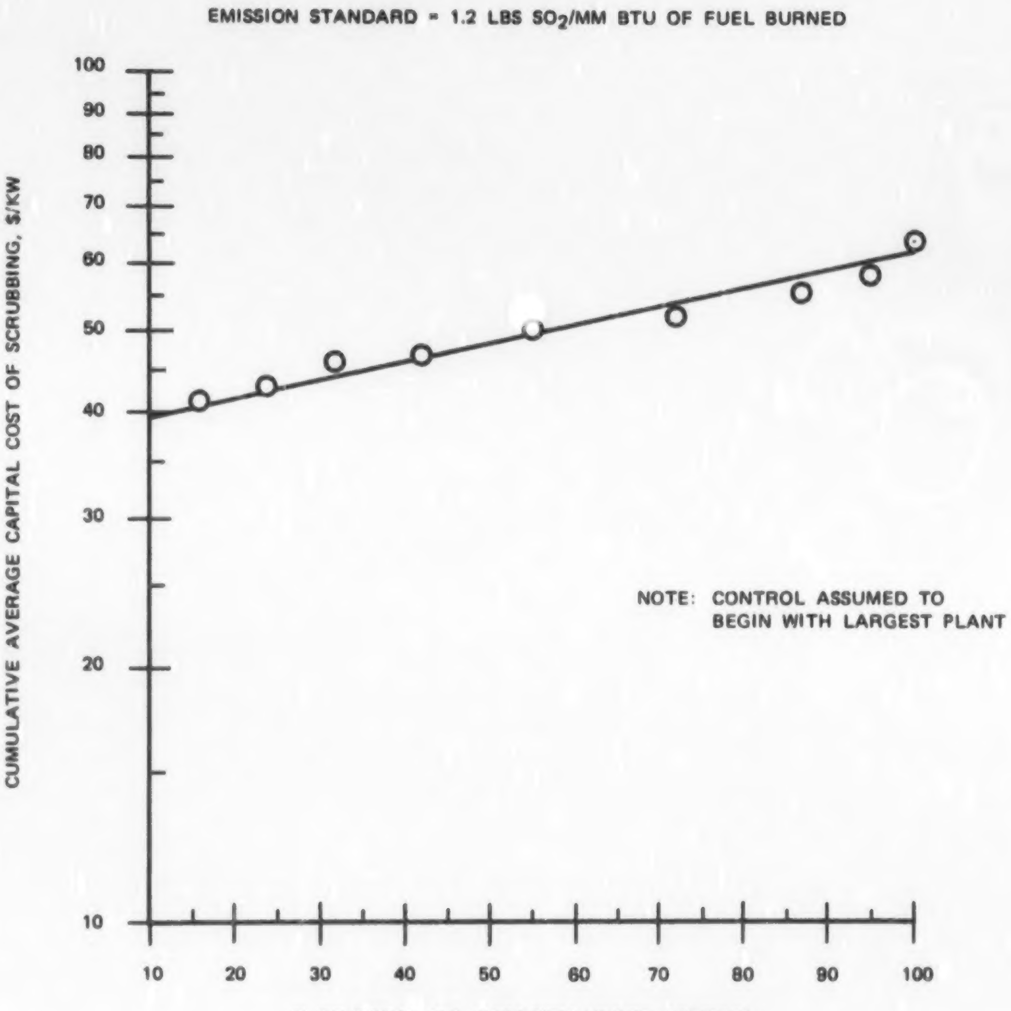


FIGURE 7.8

CUMULATIVE INCREMENTAL OPERATING COST
WET LIMESTONE SYSTEM APPLIED TO POWER PLANTS

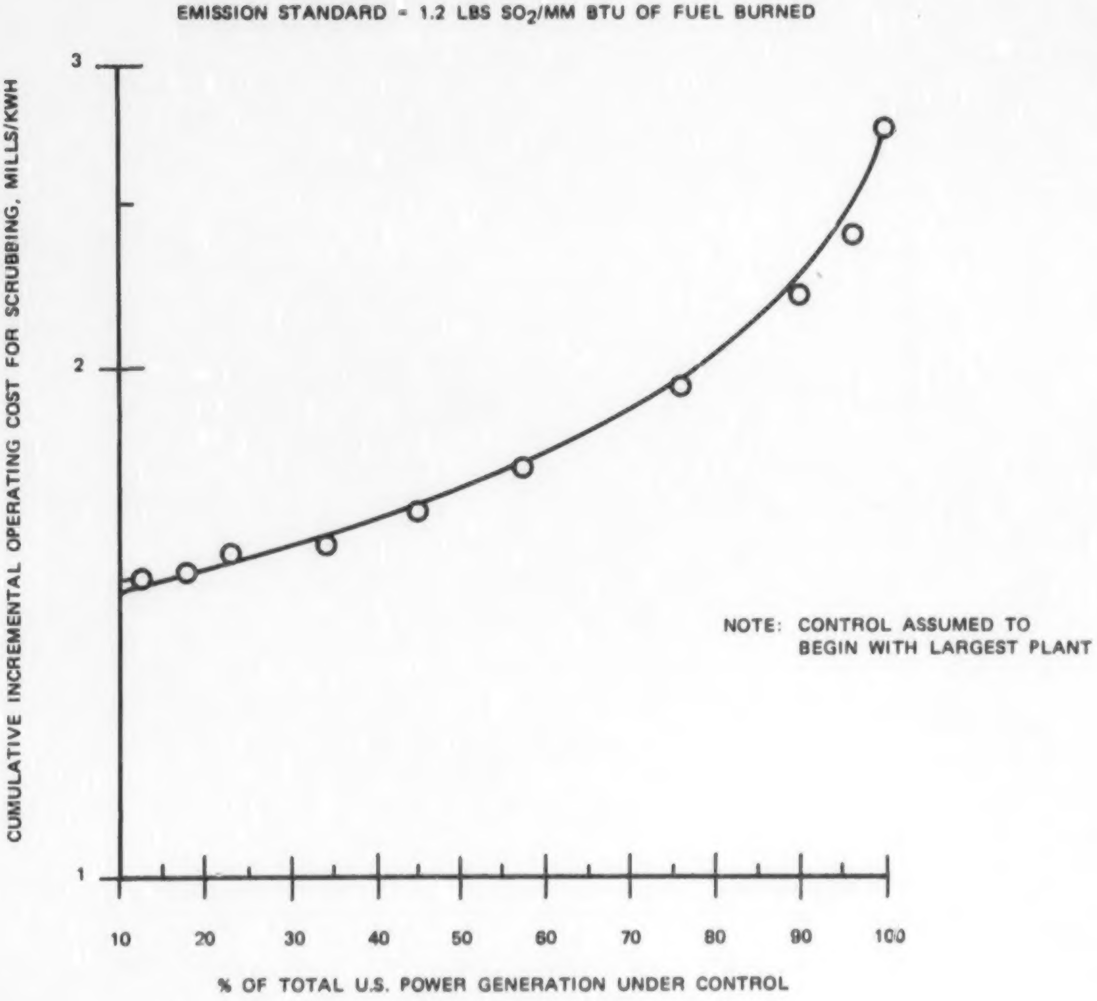


FIGURE 7.9

AVERAGE TOTAL CAPITAL REQUIREMENT FOR INSTALLING WELLMAN/ALLIED SYSTEM IN EXISTING POWER PLANTS

EMISSION STANDARD = 1.2 LBS SO_2/MM BTU OF FUEL BURNED

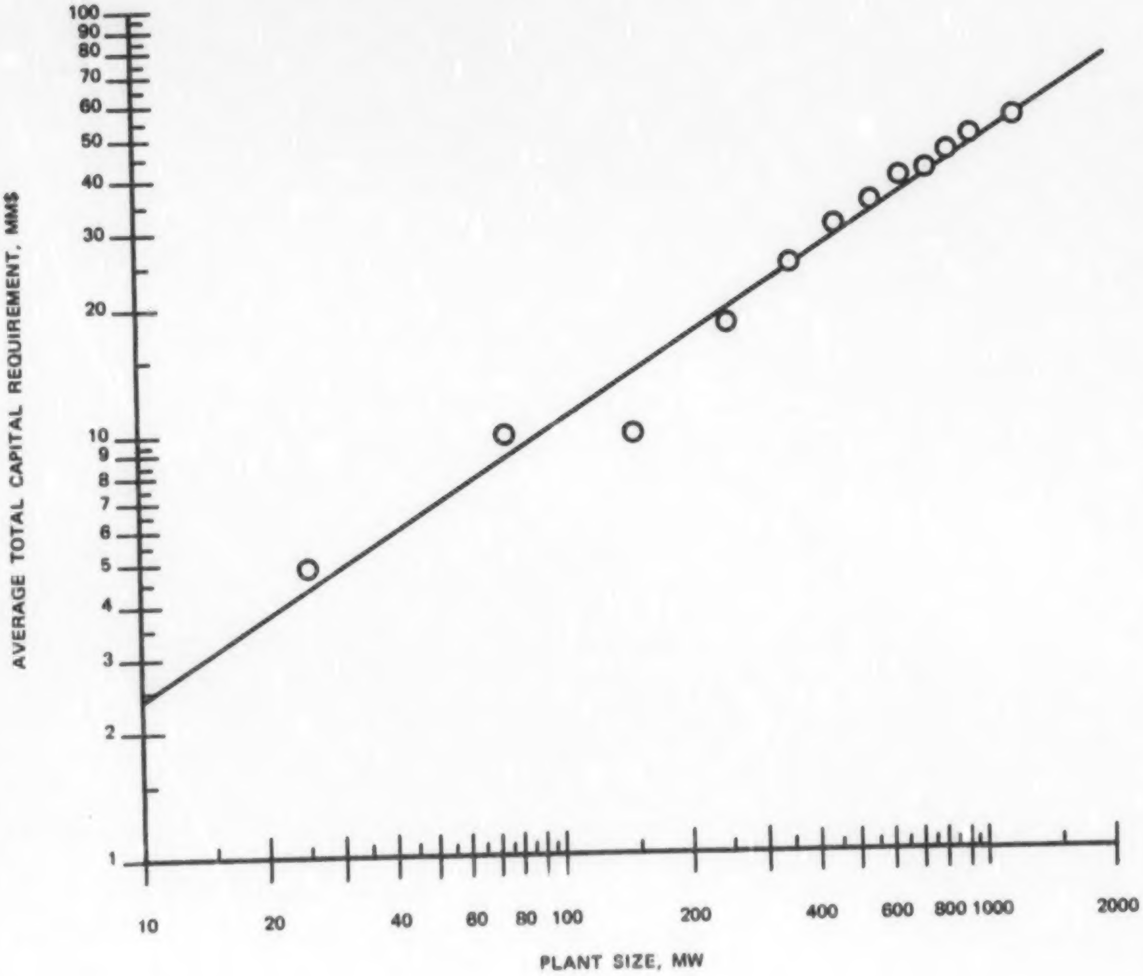


FIGURE 7.10

AVERAGE UNIT COST FOR INSTALLING WELLMAN/ALLIED SYSTEM IN EXISTING POWER PLANTS

EMISSION STANDARD = 1.2 LBS SO_2/MM BTU OF FUEL BURNED

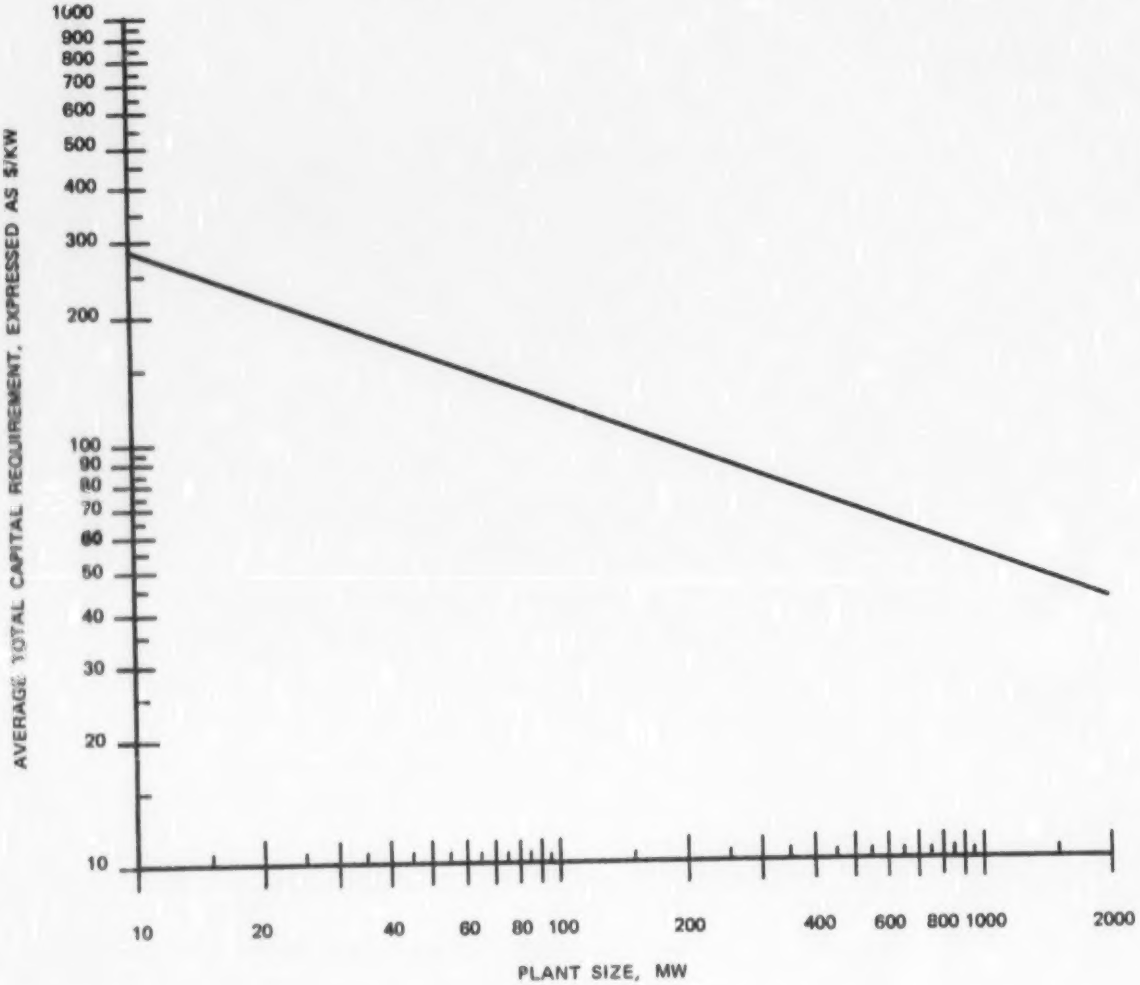


FIGURE 7.11

INCREMENTAL OPERATING COST FOR WELLMAN/ALLIED SYSTEM IN EXISTING POWER PLANTS

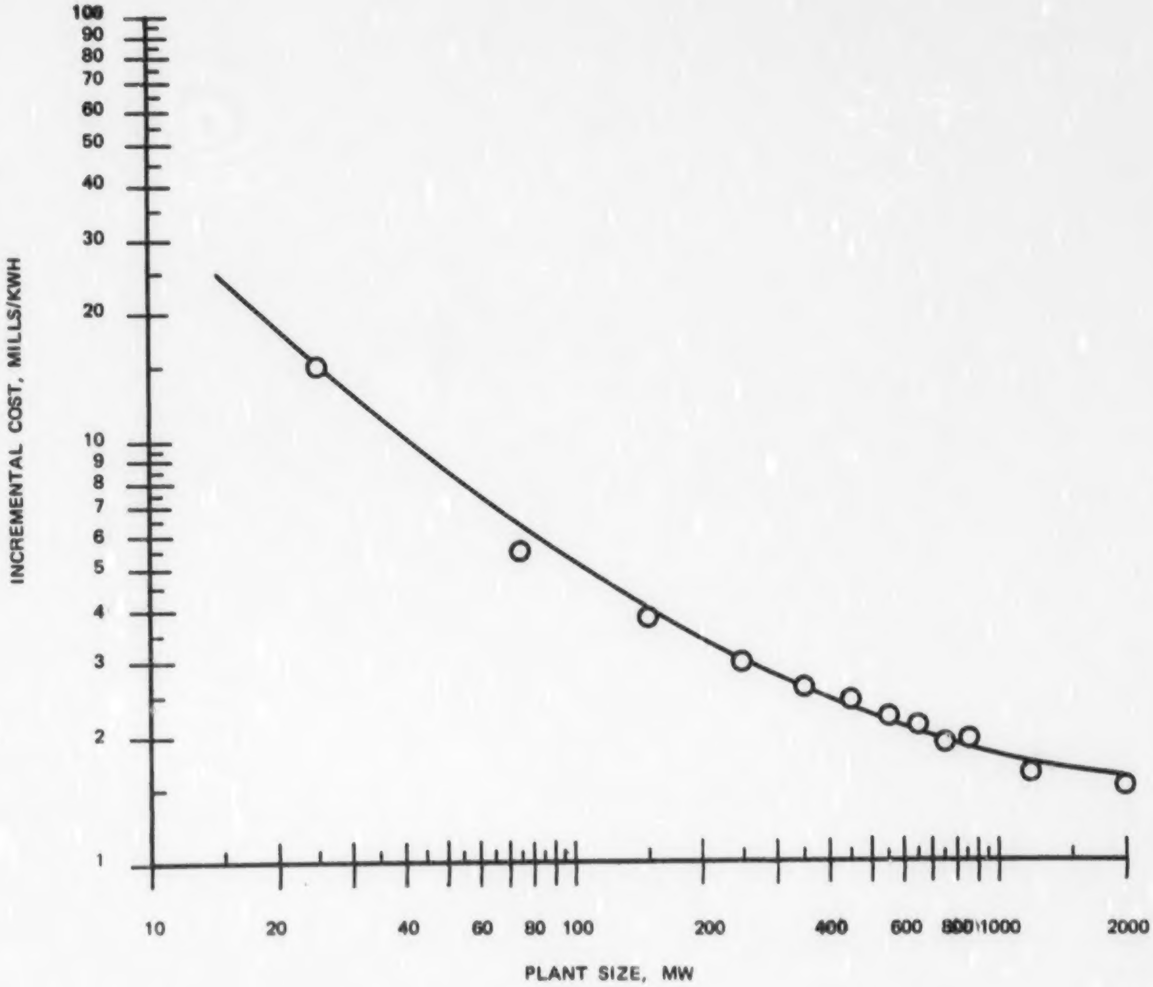


FIGURE 7.12

DEMAND FOR CLEAN FUEL AS ALTERNATIVE TO STACK GAS SCRUBBING WELLMAN/ALLIED SYSTEM APPLIED TO EXISTING POWER PLANTS

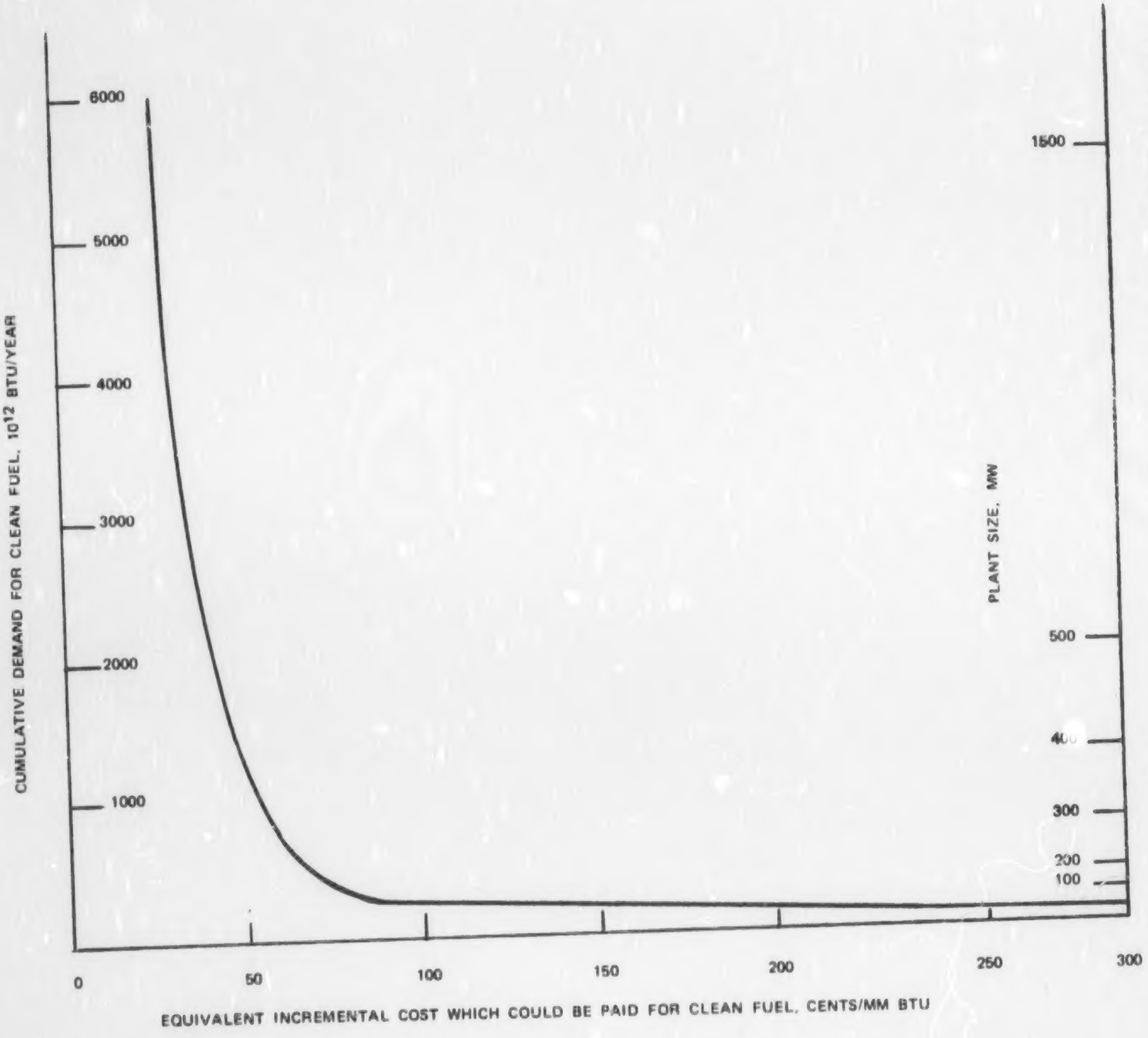
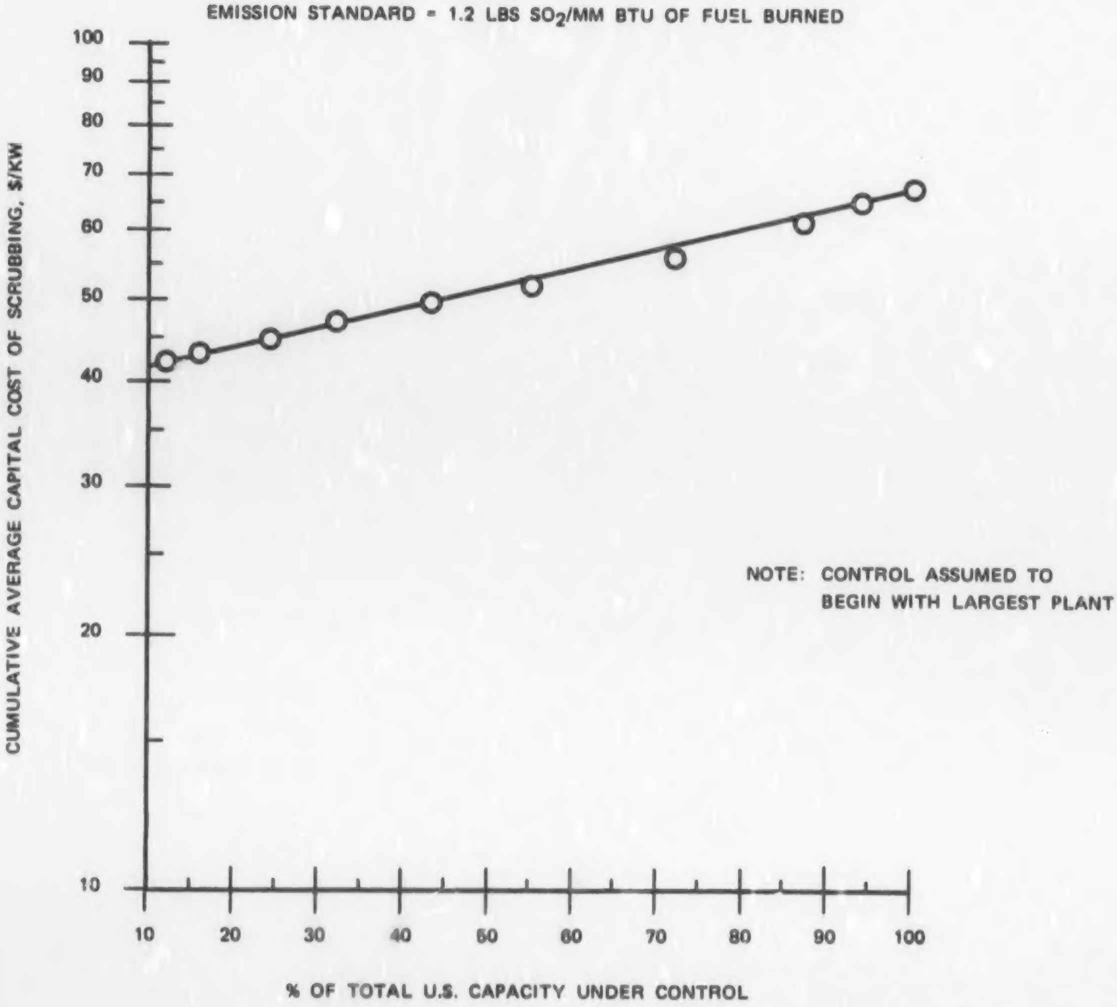


FIGURE 7.13

CUMULATIVE AVERAGE CAPITAL COST WELLMAN/ALLIED SYSTEM APPLIED TO POWER PLANTS


EMISSION STANDARD = 1.2 LBS SO_2/MM BTU OF FUEL BURNED

NOTE: CONTROL ASSUMED TO BEGIN WITH LARGEST PLANT

FIGURE 7.14

CUMULATIVE INCREMENTAL OPERATING COST WELLMAN/ALLIED SYSTEM APPLIED TO POWER PLANTS

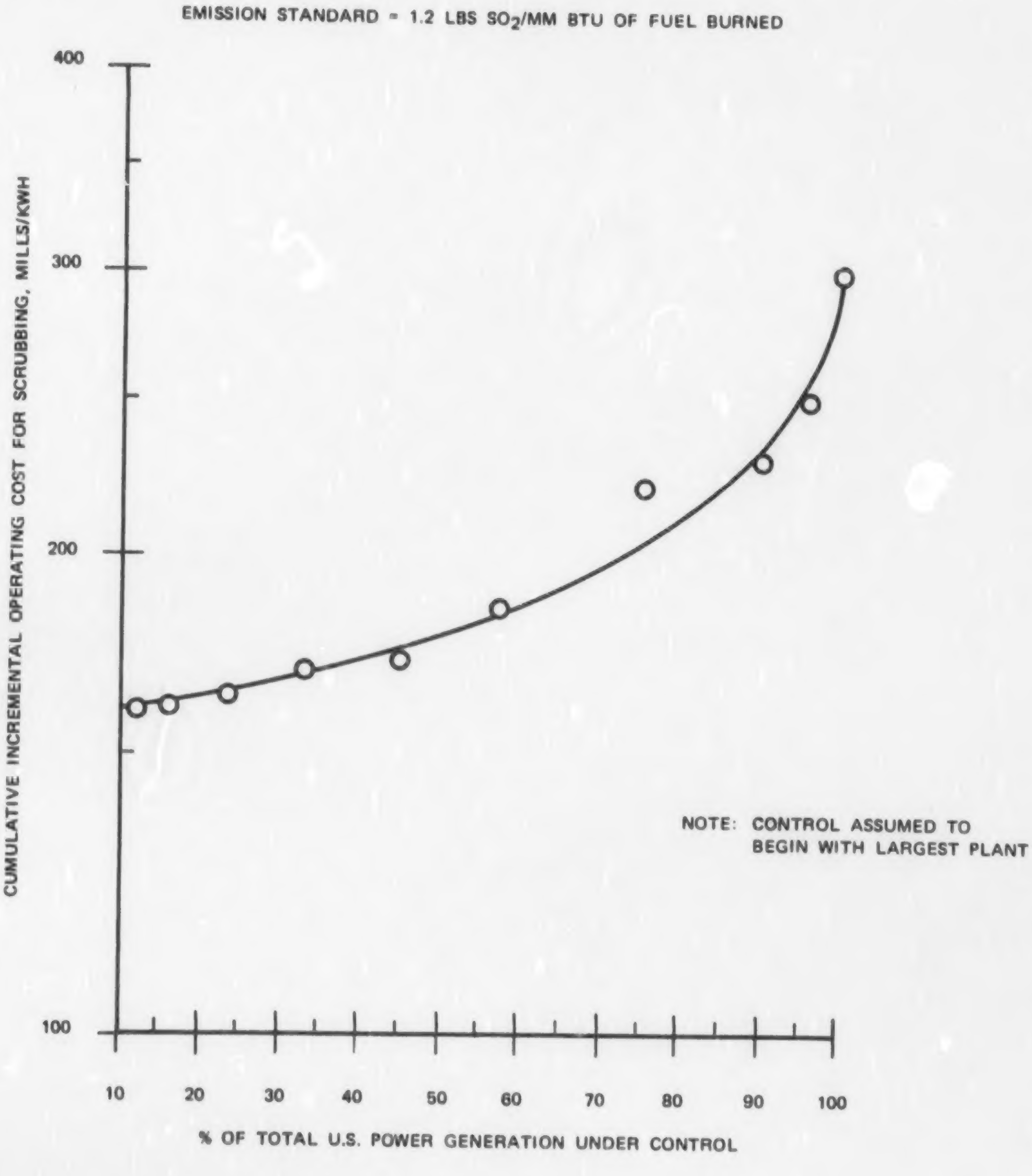
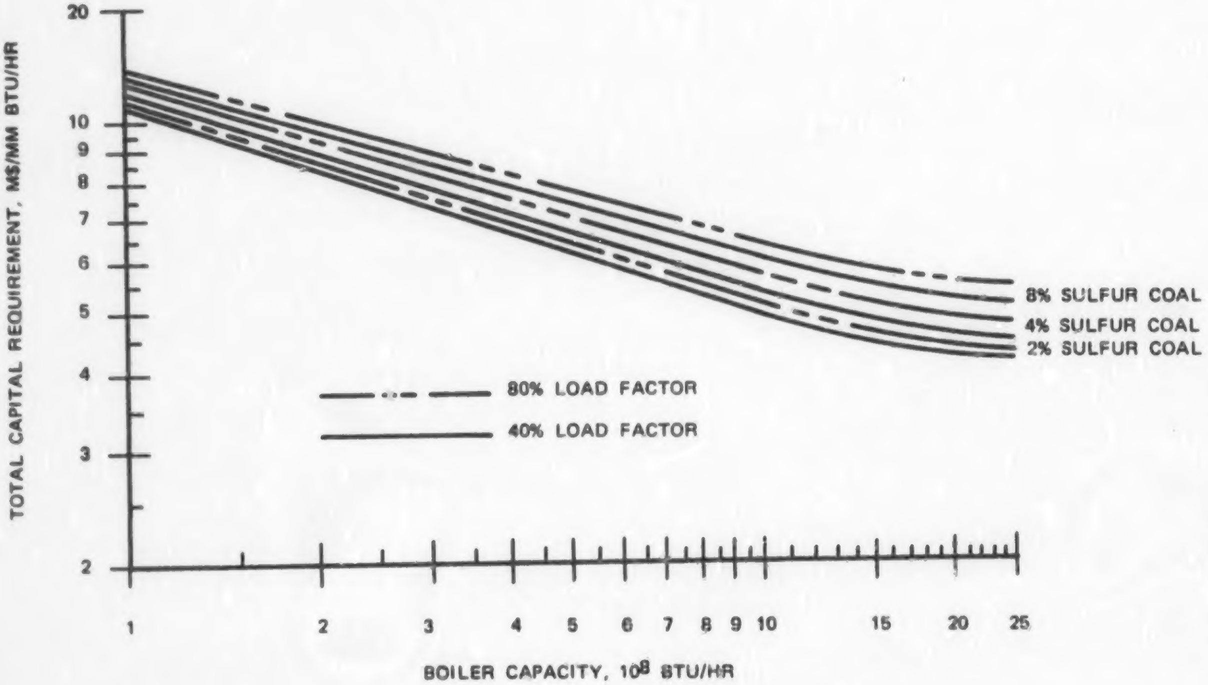


FIGURE 7.15

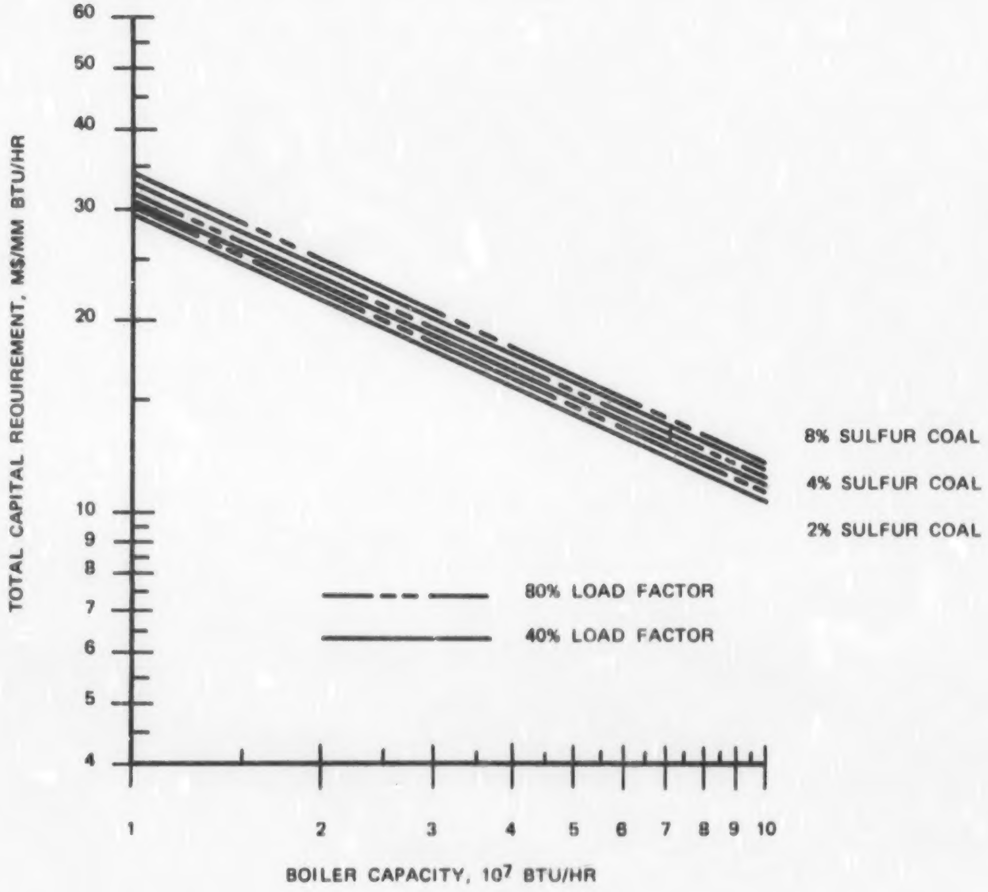
EFFECT OF BOILER CAPACITY ON TOTAL CAPITAL REQUIREMENT WET LIMESTONE PROCESS APPLIED TO LARGE INDUSTRIAL BOILERS



BASIS OF CALCULATION: NO RETROFIT, NO CONTINGENCY, U.S. GULF COAST LOCATION, END OF 1973 FIGURE, HHV OF COAL 11,000 BTU/LB, 40% EXCESS AIR WITH AN AIR LEAKAGE OF 10%.

FIGURE 7.16

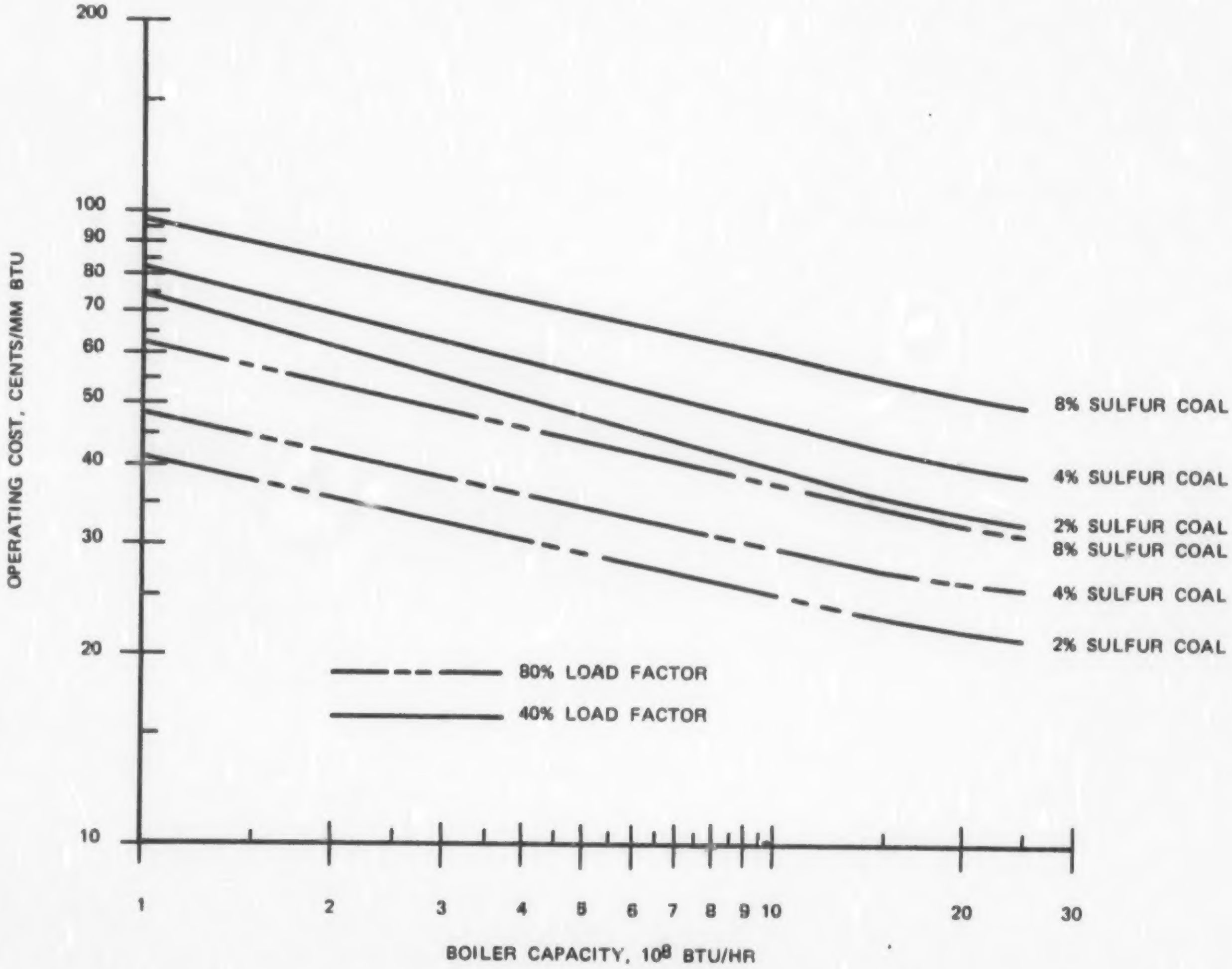
EFFECT OF BOILER CAPACITY ON TOTAL CAPITAL REQUIREMENT WET LIMESTONE PROCESS APPLIED TO SMALL INDUSTRIAL BOILERS



(SEE FIGURE 7.15 FOR BASIS OF CALCULATION.)

FIGURE 7.17

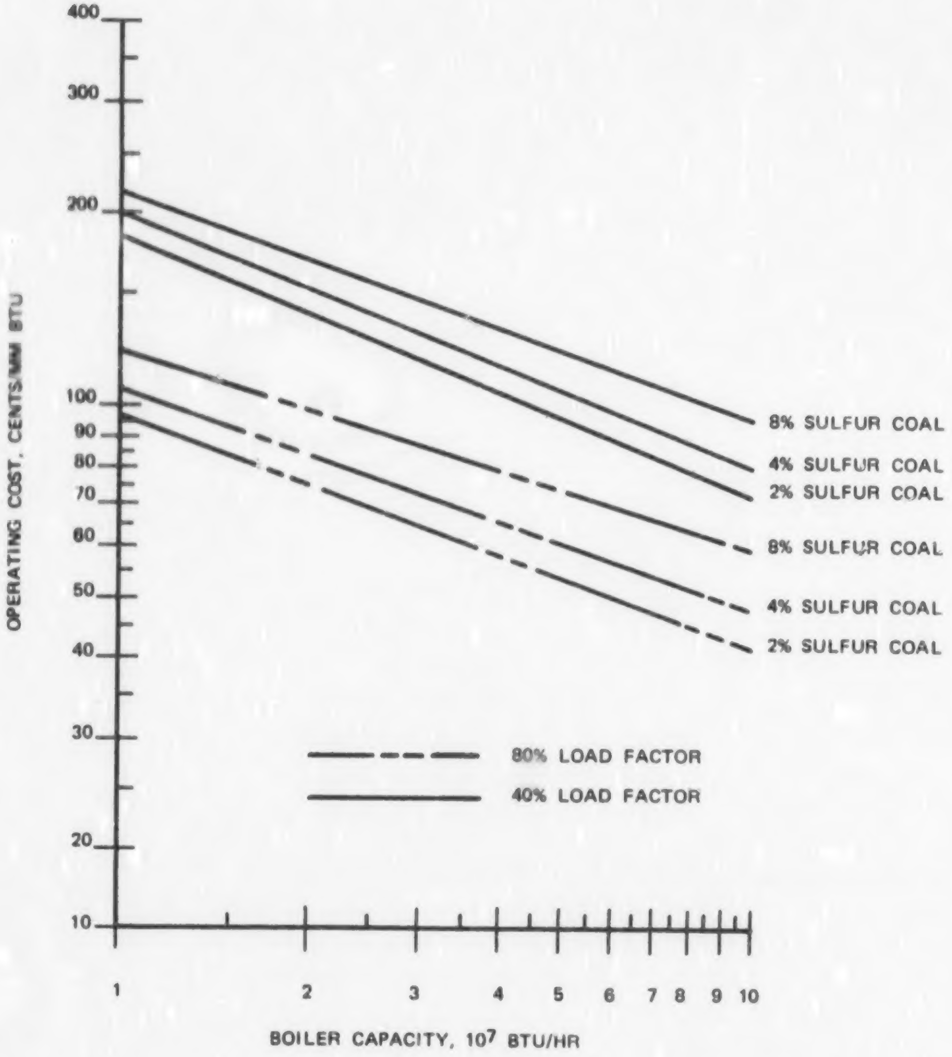
EFFECT OF BOILER CAPACITY ON OPERATING COST WET LIMESTONE PROCESS APPLIED TO LARGE INDUSTRIAL BOILERS



(SEE FIGURE 7.15 FOR BASIS OF CALCULATION.)

FIGURE 7.18

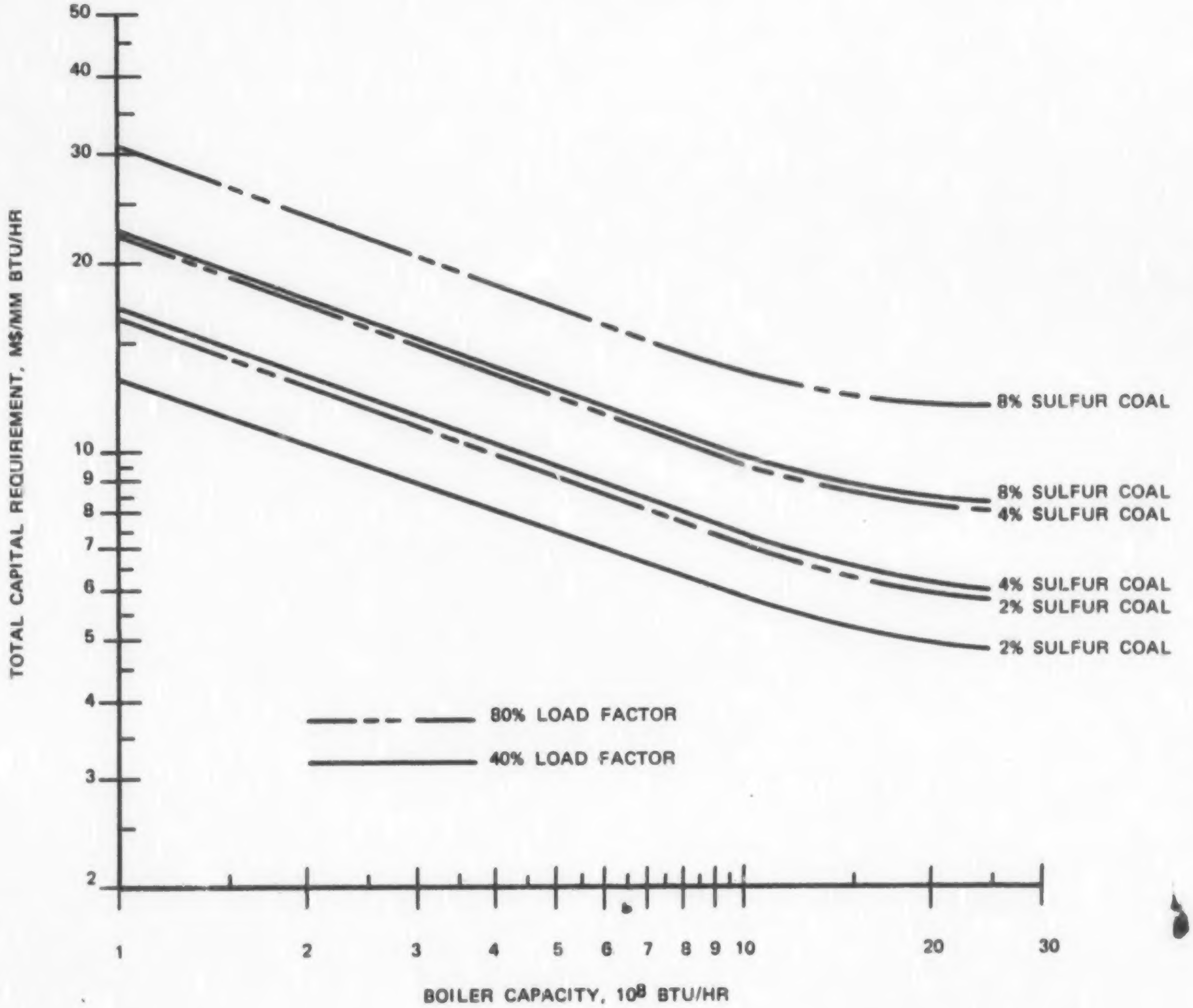
EFFECT OF BOILER CAPACITY ON OPERATING COST
WET LIMESTONE PROCESS APPLIED TO SMALL INDUSTRIAL BOILERS



(SEE FIGURE 7.15 FOR BASIS OF CALCUATION.)

FIGURE 7.19

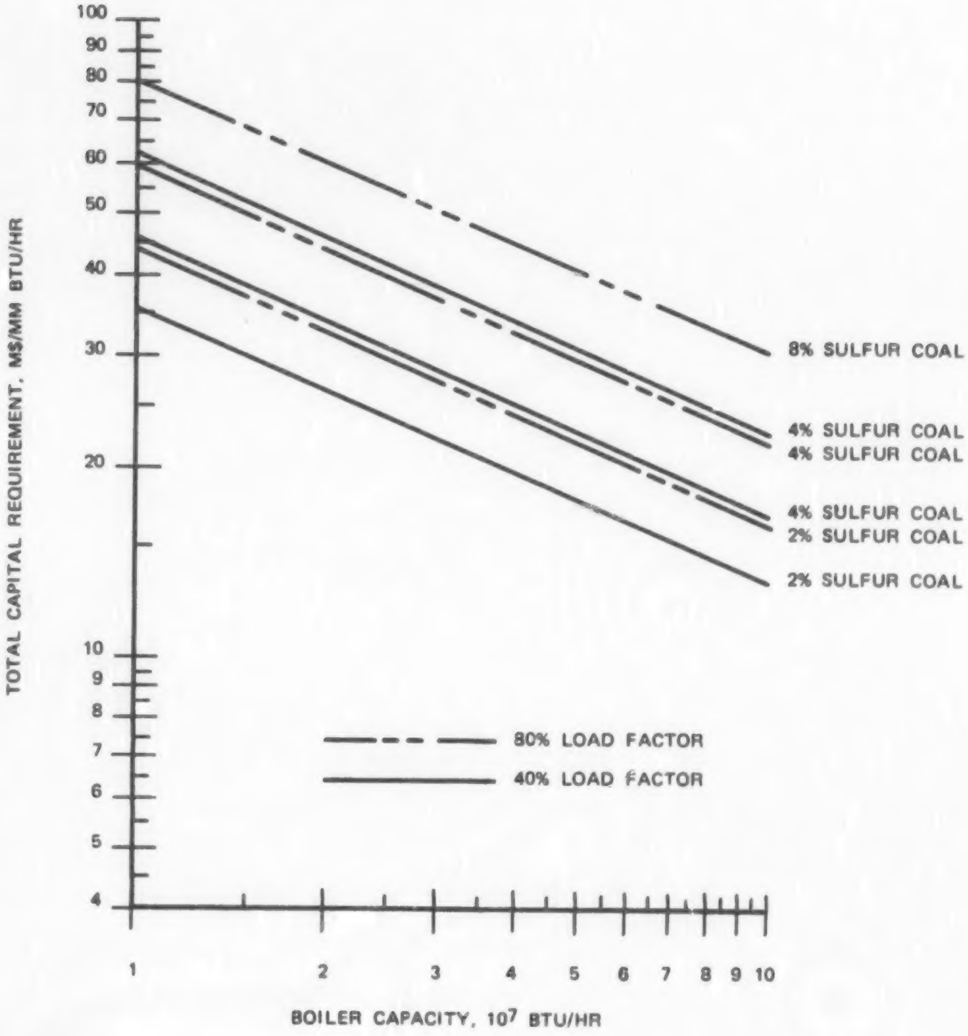
EFFECT OF BOILER CAPACITY ON TOTAL CAPITAL REQUIREMENT WELLMAN/ALLIED PROCESS APPLIED TO LARGE INDUSTRIAL BOILERS



BASIS OF CALCULATION: NO RETROFIT, NO CONTINGENCY, U.S. GULF COAST LOACTION, END OF 1973 FIGURE, HHV OF COAL 11,000 BTU/LB, 40% EXCESS AIR WITH AN AIR LEAKAGE OF 10%.

FIGURE 7.20

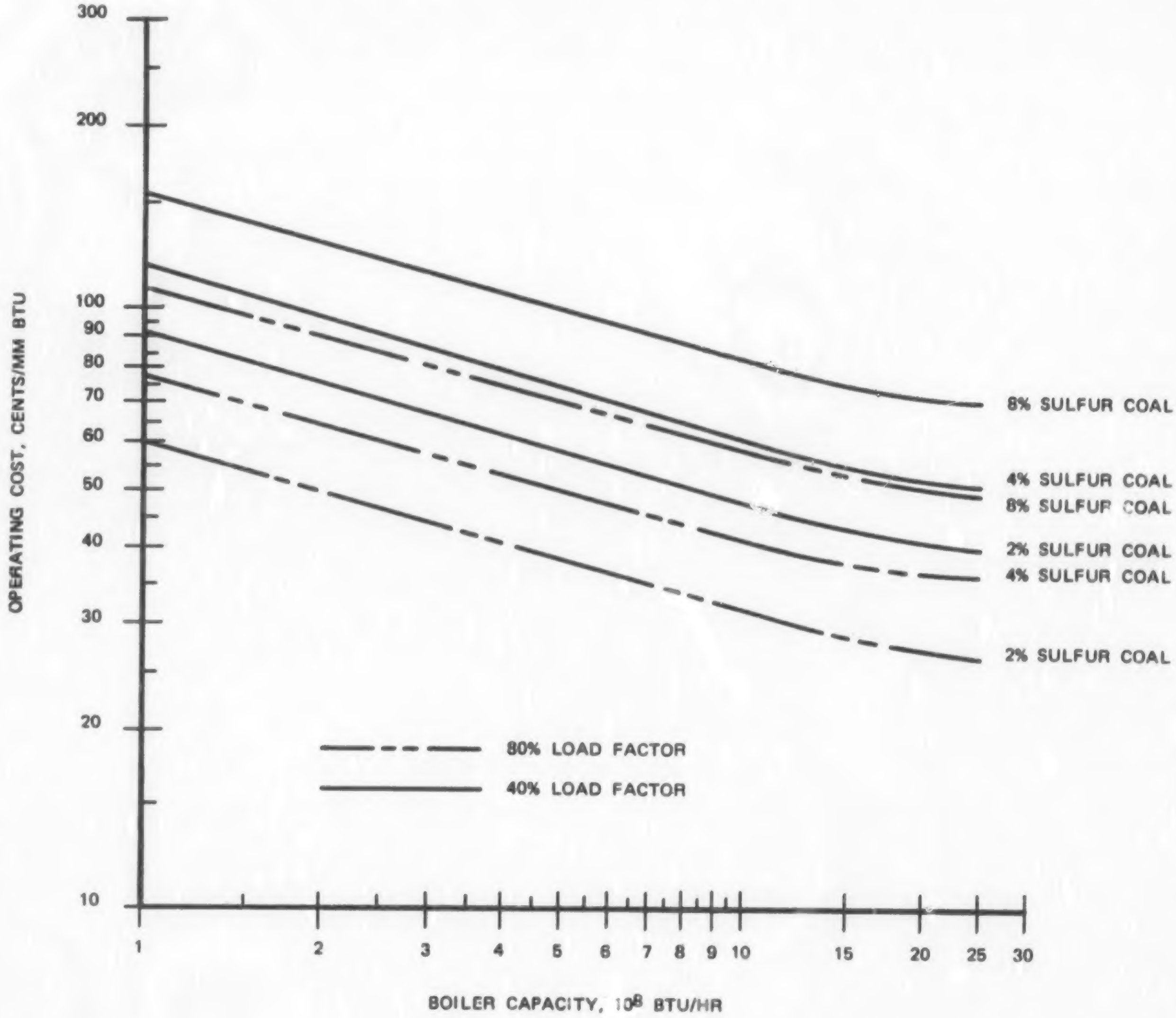
EFFECT OF BOILER CAPACITY ON TOTAL CAPITAL REQUIREMENT WELLMAN/ALLIED PROCESS APPLIED TO SMALL INDUSTRIAL BOILERS



(SEE FIGURE 7.19 FOR BASIS OF CALCULATION.)

FIGURE 7.21

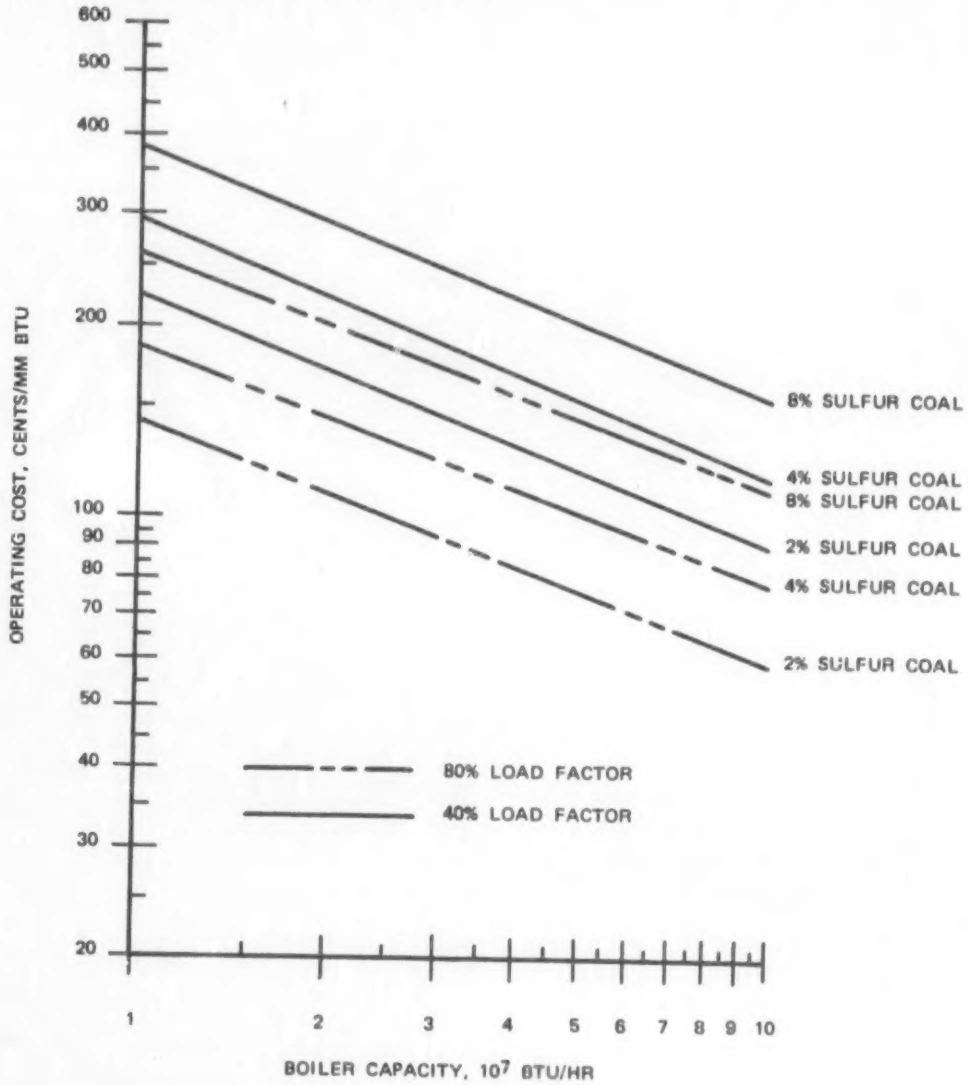
EFFECT OF BOILER CAPACITY ON OPERATING COST WELLMAN/ALLIED PROCESS APPLIED TO LARGE INDUSTRIAL BOILERS



(SEE FIGURE 7.19 FOR BASIS OF CALCULATION.)

FIGURE 7.22

EFFECT OF BOILER CAPACITY ON OPERATING COST
WELLMAN/ALLIED PROCESS APPLIED TO SMALL INDUSTRIAL BOILERS



(SEE FIGURE 7.19 FOR BASIS OF CALCULATION.)

8. SUBSTITUTE NATURAL GAS PRODUCTION USING A LURGI OXYGEN GASIFIER

8.1 Process Model

The model represents a plant producing 250 x 10^9 Btu/day of pipeline quality gas having a higher heating value of 970 Btu/SCF. This is generally considered the standard size plant. At this stage there is little economic incentive to increase this size for two good reasons: firstly, most of the equipment already consists of several parallel trains, and secondly, because of the difficulties in financing projects well in excess of $300 million.

So the only variations in the size of the units making up the plant will be produced by different types of coal feed.

8.1.1 Coal Types

There are four basic coal types: lignite, subbituminous, bituminous and anthracite. All four types have been used as feed to Lurgi gasifiers. Anthracite can be dismissed as a feedstock for SNG because of its scarcity. There are, however, large deposits of lignite, subbituminous and bituminous coal in the U.S.

These four coal types can best be categorized by the dry ash free carbon content. Here, dry means free from all water, not only surface moisture.

	% carbon in dry ash free coal (PCARB)
lignite	65 - 73
subbituminous	73 - 77
bituminous	77 - 91
anthracite	91 - 96

A mid-range composition for each type is:

% Dry Ash free basis (DAFB)	Lignite	Subbituminous	Bituminous	Anthracite
C	69	75	84	93
H	5	5	5.5	3
O	24	18	7	2
N	1	1	1.5	1
S	1	1	2	1
	100	100	100	100

These coals, when received from the mine, will have the following range of water and ash content.

% of coal as received	Lignite	Subbituminous	Bituminous	Anthracite
Water (PH2O)	28-40	16-28	4-16	2-4
Ash (PASH)	4- 8	5-12	6-14	6-18

Both the water and the ash content can vary a few percent up or down throughout the same coal field.

Figure 8.1 gives the higher heating values in Btu/lb of dry ash free coal against the percentage carbon on a dry ash free basis. This graph has been derived from information on several coals of each type (23, 24, 25).

The sulfur contents of the lignites and subbituminous coals are generally low, about 1% d.a.f.b.; however, the sulfur contents of the bituminous coals can be as high as 8% d.a.f.b., although they are usually less than 4% d.a.f.b.

8.1.2 Coal, Oxygen and Steam Requirements for the SNG Plant

Figure 8.2 shows the dry ash free coal, oxygen and steam requirements in million lb/hr for a Lurgi oxygen gasification plant producing 250 x 10^9 Btu/day of pipeline quality gas of higher heating value 970 Btu/SCF. These graphs have been produced from available designs and published information (26, 27, 28). They are, of course, simplified linear representations of the real situation. The difference between the coal feed to the gasifier and the total coal requirement of the plant is the fuel to the furnace producing HP steam and power.

8.1.3 Electric Power and High Pressure Steam Requirements for the SNG Plant

The following bases were used to establish the electric power and high pressure steam requirements:

1. In this model no major drivers are powered by electricity. The total electric power requirement for instruments, small drivers not powered by steam and lighting is about 68 megawatts.

2. The SNG compressor is powered by low pressure steam generated by the gasifier and waste heat recovery. This also covers all other low pressure process steam requirements and a few small drivers.

3. The air and oxygen compressors are powered by HP steam with vacuum condensation (1100 psig, 825°F→3 psia requires 10 lb/hr of steam/KWH). A plant producing 0.5 million lb/hr of oxygen requires 116 megawatts.

4. The Lurgi process units require 0.3 million lb/hr of HP steam.

5. The methanator generates about 1.4 million lb/hr of HP steam.

6. The 550 psig steam for the gasifier is provided by expanding 1100 psig steam and generating some of the 68 megawatts power requirement (50 lb/hr of steam/KWH).

7. The rest of the 68 megawatt power is provided by expanding the 1100 psig steam to 3 psia.

8.1.4 Sample Calculation of Plant Total Coal Requirement

The total coal feed to the SNG plant is calculated below for two coals: a mid-range lignite (PCARB = 69%), and a mid-range bituminous (PCARB = 84%).

	LIGNITE	BITUMINOUS
	megawatts	megawatts
Total electric power required	68	68
Power Generated by expanding gasifier steam	39	50
Net power required	28	18
	million lb/hr	million lb/hr
HP steam required to generate net power	0.28	0.18
HP steam for expansion to gasifier	1.94	2.48
HP steam required for air and oxygen compressors	0.84	1.39
HP steam required for Lurgi process units	0.30	0.30
Total HP steam requirement	3.36	4.35
HP steam generated by methanator	1.40	1.40
Net HP steam requirement	1.96	2.95
	million Btu/hr	million Btu/hr
Furnace Duty	2520	3790
Furnace Liberation	2900	4360
Heat provided by burning tar, etc. (8% of heat input for lignite and 10% of heat input for bituminous)	1180	1500
Heat provided by coal	1720	2860

	DAF million lb/hr	DAF million lb/hr
Coal required to provide this heat	0.15	0.19
Coal feed to furnace including 12% extra for stack gas scrubbing	0.17	0.21
Coal feed to gasifier	1.28	1.01
Total coal requirement of the plant	1.45	1.22

The total dry ash free coal requirement of a 250 x 10^9 Btu/day SNG plant is given by:

$$\underline{\underline{TDAFC = 1.51 - 0.0156\ (PCARB-65)}}\ \text{million lb/hr}$$

The total "as received coal" requirement of a 250 x 10^9 Btu/day SNG plant is given by:

$$\underline{\underline{TCOAL = 100\ TDAFC/(100-PH2O-PASH)}}\ \text{million lb/hr}$$

8.2 Cost Model

8.2.1 Major Equipment Costs, E

The SNG plant has been considered as 12 units (Fig. 8.3).

Section Number	Solid Handling or Chemical Processing	Unit
1	S	Coal Preparation and Handling
2	S	Fines Agglomeration
3	S	Coal Gasification
4	C	Shift Conversion and Gas Cooling
5	C	Gas Purification by the Rectisol Process
6	C	Methane Synthesis
7	C	SNG Compression
8	C	The Oxygen Plant
9	C	The Phenosolvan Unit
10	C	Furnace Stack Gas Scrubbing and Plant Sulfur Recovery
11	C	Utility Plant
12	C	Other Offsites

The following equipment costs were developed using in-house and published data (26, 27), updated to the end of 1973 and adjusted to the U.S. Gulf Coast basis. The standard relationships given in the General Cost Model between major equipment costs, E, other material costs, M, and direct construction labor costs, Gulf Coast, L, were used in cross-checking the available information.

Section 1 - Coal Preparation and Handling

Raw coal from storage is crushed and classified in this section. The larger size fraction (about 70%) is sent to the gasifiers. Some of the fines generated during crushing are burned in the furnace, the remainder are sent to the fines agglomeration unit. No costs are included for equipment delivering coal from the mine. Coal is assumed to be delivered to the plant storage and the delivery costs included in the cost of the coal. It is also assumed that the ash is removed back to the mine and the cost of this disposal is included in the cost of the coal. The plant requires more lignite feed than bituminous coal feed, however, the lignite crushes more easily. Therefore, it has been assumed that there is no variation of E with coal type.

$$E1 = 2{,}100 \text{ M\$}$$

Section 2 - Fines Agglomeration

Variations in coal feedrate to the plant and to the furnace mean the coal flow to the fines agglomeration unit decreases as the carbon content of the coal increases. These quantities were determined, the equipment cost variation calculated as the 0.6 power of the size and the cost simplified to a linear equation.

$$D2 = 5{,}000 - 100\ (\text{PCARB}-65) \text{ M\$}$$

Section 3 - Coal Gasification

The number of gasifiers required depends on the quantity of the coal feed, the slagging properties of the coal and the reactivity of the coal. Although the coal feed to the gasifier decreases with the increase in carbon content so does the coal reactivity, the net effect is that more gasifiers are required for the highest rank coals. This, of course, is a complicated effect, which has been simplified as best possible in the following cost equation.

$$E3 = 14,800 + 160\ (PCARB-65)\ M\$$$

Section 4 - Shift Conversion and Gas Cooling

In this section, the H_2/CO ratio of the crude gas is adjusted by the shift reaction:

$$CO + H_2O \rightarrow CO_2 + H_2$$

to 3.0, which is the stoichiometric ratio for methanation. The crude gas is cooled before the purification unit. No significant cost variations could be determined with carbon content.

$$E4 = 4,500\ M\$$$

Section 5 - Gas Purification by the Rectisol Process

This unit removes CO_2, H_2S and naphtha from the gas before methanation. There is a small increase in cost as the sulfur content of the coal increases. This variation has been expressed as a linear equation.

$$E5 = 13,000 + 200\ PSULF\ M\$$$

PSULF is the percent sulfur in the dry, ash free coal.

Section 6 - Methane Synthesis

In this section CO and H_2 in the treated gas are converted to CH_4 by the reaction:

$$CO + 3H_2 \rightarrow CH_4 + H_2O$$

The final CO_2 absorption is also included in this section. No significant variation in cost could be determined with carbon content.

E6 = 5,500 M$

Section 7 - SNG Compression

This section is SNG compression for delivery to the pipeline. The costs include the compressors, steam turbine drivers and the vacuum condensers.

E7 = 3,000 M$

Section 8 - The Oxygen Plant

This section produces the oxygen feed to the gasifiers. The oxygen requirements of the plant increase with increasing carbon content of the coal. The cost variations have been determined as a function of the 0.6 power of size and expressed as a linear equation.

E8 = 9,700 + 160 (PCARB-65) M$

Section 9 - The Phenosolvan Unit

The unit handles all the gas liquor which has been condensed. Here the main objective is to remove the water before the phenols and tars can be routed to the furnace. No significant cost variations can be determined in general terms.

$$E9 = 1,800 \text{ M\$}$$

Section 10 - Furnace Stack Gas Scrubbing and Plant Sulfur Recovery

The section includes a stack gas scrubbing and SO_2 regeneration unit on the furnace if this is required. It contains the sulfur recovery unit for the whole plant. It has been assumed that 80% of the sulfur entering the plant emerges as sulfur by-product. An equation has been derived which contains a term for the variation of stack gas scrubbing costs. These were computed as the 0.6 power of size and expressed as a linear function. This term is effectivley zero when the coal contains little sulfur. The equation also has a term for the sulfur recovery unit. Even a coal with a sulfur content of 0.1% requires a small sulfur recovery unit.

$$E10 = 1,250 \text{ PSULF} + 1065 \text{ (TDAFC.PSULF)}^{0.6} - 250 \text{ M\$}$$

Section 11 - The Utility Plant

The utility plant supplies the power and HP steam for the plant. It is made up of 3 areas, the boiler plant, the power plant and processing of the fuel gas and tar.

The boiler plant increases in size as the carbon content of the coal increases. The variations in cost were computed as the 0.8 power of size and expressed as a linear function. The rest of the unit was assumed to be independent of coal type and has a major equipment cost of $4,500 M.

$$E11 = 13,800 + 200 \text{ (PCARB-65) M\$}$$

Section 12 - Other Offsites

Offsites other than the utility plant have been grouped together as one section. The major items are storage facilities, service systems, electrical distribution, sewers and waste disposal, site preparation, plant buildings and mobile equipment.

No meaningful variation with coal type could be derived.

$$E12 = 14{,}000 \text{ M\$}$$

8.2.2 Total Net Annual Operating Cost

The total net annual operating cost, AOC, is the total cost of operating the plant less the credits from the sale of by-products. It does not include return of capital, payment of interest on debt or income tax on equity returns.

This model conforms exactly to the format in the General Cost Model, which is fully explained in Section 4.

The total net annual operating cost is, therefore, given by:

$$\underline{\underline{AOC = 0.078\ TPI + 2\ TO \cdot CO\ (1 + F) + ANR}}$$

The total number of shift operators for the SNG plant can be assumed to be 300.

The annual cost of raw materials less by-product credits has been simplified and is given by -

$$\underline{\underline{ANR = ACOAL + ACHEM - ASULF}}$$

The annual cost of catalysts and chemicals, ACHEM, is assumed constant -

	Annual Cost M$
Shift catalyst	40
Methanator catalyst	60
Methanol	500
Isopropyl ether	200
H_2SO_4	120
NaOH	400
Activated Carbon	50
Lime	10
Na_2CO_3	20
Process Water	200

ACHEM = 1,600 M$

The annual cost of the coal feed to the plant is given by:

$$\text{ACOAL} = \frac{\text{CCOAL} \cdot \text{TCOAL} \times 24 \times \text{SD}}{2{,}000 \times 1{,}000} \quad \text{M\$/Yr.}$$

where CCOAL is the unit cost of coal as received at the site in $/ton and SD is the number of days the plant is on stream per year.

The equation reduces to:

$$\text{ACOAL} = 12\ \text{CCOAL} \cdot \text{TCOAL} \cdot \text{SD} \quad \text{M\$/Yr.}$$

The credit per year for the sale of sulfur, ASULF, is given by:

$$\text{ASULF} = \frac{\text{CSULF} \times 0.8 \times \text{TDAFC} \times \text{PSULF} \times 24 \times \text{SD}}{2{,}000 \times 100 \times 1{,}000} \quad \text{M\$/Yr.}$$

where CSULF is the unit credit for sulfur in \$/ton. It has been assumed that 80% of the sulfur in the coal feed to the plant is recovered. The equation reduces to

$$\underline{ASULF = 0.1\ CSULF\ .\ TDAFC\ .\ PSULF\ .\ SD}\quad M\$/Yr.$$

8.2.3 Total Plant Investment, Total Capital Required and Total Annual Production Cost

This model conforms exactly to the General Cost Model and so the Total Plant Investment, TPI, at different locations can be derived from the graph of C vs. F in Section 4.

$$\underline{TPI = C \cdot E}$$

Sections 1, 2 and 3 of the SNG plant are classified as solids handling and the remaining 9 sections as chemical processing.

The Total Capital Required, TCR, is given by the TCR equation in the General Cost Model.

$$\underline{TCR = 1.21\ TPI + 0.8\ TO \cdot CO\ (1 + F) + 0.4\ ANR}$$

The Total Annual Production Cost, TAC, is also obtained from the General Cost Model.

$$\underline{TAC = 0.225\ TPI + 2.1\ TO \cdot CO\ (1 + F) + 1.04\ ANR}$$

8.2.4 Calculation of Costs for Three Types of Coal in Three Different Locations

Example 1

Location: New Mexico F = 1.3

Coal Details: Subbituminous PCARB = 77%, PSULF = 1%
PH20 = 17%, PASH = 17%
CCOAL = $3/ton

Other Information:

The plant is on stream for 93% of the year, SD = 340 days. The by-product sulfur credit CSULF = $5/ton. The Gulf Coast Operating labor costs CO = $7/hour.

Derived Information:

Scale up factor to give TPI obtained from General Cost Model.
C = 2.63 solid handling, Sections 1 to 3
C = 3.56 Chemical handling, Section 4 to 12
TDAFC = 1.32 million lb/hr, Figure 8.2

Using the major equipment equations shown in Table 8.1 and the above values of C, the following costs were calculated:

Section	E M$	TPI M$
1	2,100	5,523
2	3,800	9,994
3	16,720	43,974
4	4,500	16,020
5	13,190	46,956
6	5,500	19,580
7	3,000	10,680
8	11,620	41,367
9	1,800	6,408
10	2,157	7,679
11	16,200	57,672
12	14,000	49,840
		315,693

TPI = 315.693 million

TCOAL = 1.32/0.66 = 2.0 million lb/hr
ACOAL = 12 x 3 x 2 x 340 = M$ 24,480
ASULF = 0.1 x 5 x 1.32 x 1 x 340 = M$ 224

ANR = 24,480 + 1,600 - 224
= M$ 25,856

TCR = 1.21 x 315,693 + 0.8 x 300 x 7 (1 + 1.3) + 0.4 x 25,856
TCR = $396.195 million

TAC = 0.225 x 315,693 + 2.1 x 300 x 7 (1+1.3) + 1.04 x 25,856
TAC = $108.064 million

The Annual Gas Production
AGP = 250,000 x 340 = 85.0 million MMBtu/year

The gas cost = 108.064/85.0
= $1.27/MMBtu

Example 2

Location: Wyoming F = 1.3

Coal Details: Subbituminous PCARB = 74%, PSULF 0.12%
PH20 = 30%, PASH = 5%
CCOAL= $3/ton

Other Information:

SD = 340 CSULF = $5/ton CO = $7/hour

Derived Information:

C = 2.63 Sections 1 to 3
C = 3.56 Sections 4 to 12
TDAFC = 1.37 million lb/hr

Section	EM$	TPI M$
1	2,100	5,523
2	4,100	10,783
3	16,240	42,711
4	4,500	16,020
5	13,024	46,365
6	5,500	19,580
7	3,000	10,680
8	11,140	39,658
9	1,800	6,408
10	260	926
11	15,600	55,536
12	14,000	49,840
		304,030

TPI = $304.03 million

TCOAL = 1.37/0.65 = 2.1 million lb/hr
ACOAL = 12 x 3 x 2.1 x 340 = M$ 25,704
ASULF = 0
ANR = 25,704 + 1,600
= M$ 27,304

$$TCR = 1.21 \times 304{,}030 + 0.8 \times 300 \times 7\ (1+1.3) + 0.4 \times 27{,}304$$

$$\underline{TCR = \$382.66 \text{ million}}$$

$$TAC = 0.225 \times 304{,}030 + 2.1 \times 300 \times 7\ (1 + 1.3) + 1.04 \times 27{,}304$$

$$\underline{TAC = \$106.95 \text{ million}}$$

$AGP = 85.0$ million MMBtu/year

The gas cost = 106.95/85.0
= $1.26/MMBtu

Example 3

Location: Illinois F = 1.7

Coal Details: Bituminous PCARB = 78%, PSULF = 5.6%
PH20 = 14% , PASH = 15%
CCOAL = $6/ton

Other Information:

SD = 340 CSULF = $5/ton CO = $7/ hour

Derived Information:

C = 2.88 Sections 1 to 3
C = 3.95 Sections 4 to 12

Section	E M$	TPI M$
1	2,100	6,408
2	3,700	10,656
3	16,880	48,614
4	4,500	17,775
5	14,120	55,774
6	5,500	21,725
7	3,000	11,850
8	11,780	46,531
9	1,800	7,110
10	10,270	40,566
11	16,400	64,780
12	14,000	55,300
		386,729

TPI = $386.729 million

TCOAL = 1.31/0.71 = 1.85 million lb/hr

ACOAL = 12 x 6 x 1.85 x 340 = M$ 45,288

ASULF = 0.1 x 5 x 1.31 x 5.6 x 340 = M$1,247

ANR = 45,288 + 1,600 - 1,247

= M$ 45,641

TCR = 1.21 x 386,729 + 0.8 x 300 x 7 (1 + 1.7) + 0.4 x 45,641

TCR = $490.73 million

TAC = 0.225 x 386,729 + 2.1 x 300 x 7 (1 + 1.7) + 1.04 x 45,641

TAC = $146.39 million

AGP = 85.0 million MMBtu/year

Gas Cost = 146.39/85.0

= $1.72/MMBtu

8.2.5 The Influence of Coal Type, Coal Cost, Percentage Sulfur and Plant Location on Gas Cost

The graphs on the following pages were derived using the model to investigate the influence of coal type, coal cost, percentage sulfur and plant location on the cost of the SNG. The following general observations can be made after examining these figures:

1. The location of the plant and the coal cost have the largest effect on the gas cost. If the plant is built in a high construction cost area or the coal price is high, SNG costs can become unnecessarily high.

2. The sulfur content of the coal has a secondary effect. The difference in gas cost between a high and a low sulfur coal is 8 to 12 ¢/MMBtu depending on the plant location.

3. For a given location, sulfur content and coal cost, the cost of gas decreases as the percentage carbon in the coal increases, i.e., gas cost is less for bituminous coal than lignite.

4. All of the curves shown in the four figures do not represent possible real situations. For example, it is highly unlikely that low sulfur bituminous coal will be available at $3/ton and even more unlikely that it would be available in an area with a location factor around 1.0.

5. The most attractive real situations appear low sulfur subbituminous coal in areas like New Mexico and Wyoming where there is a possibility that coal could be purchased for around $3 or 4/ton including re-landscaping strip mines. Here a gas price of between $1.2 and 1.4/MMBtu (1973 plant costs) appears realistic.

6. An underestimation of 10% in the total plant investment would mean an underestimation of the gas cost by 8 to 11 ¢/MMBtu depending on the value of TPI.

TABLE 8.1

Summary of Major Equipment Cost Equations
Substitute Natural Gas Production

M$

E1 = 2,100

E2 = 5,000 - 100 (PCARB-65)

E3 = 14,800 + 160 (PCARB-65)

E4 = 4,500

E5 = 13,000 + 200 PSULF

E6 = 5,500

E7 = 3,000

E8 = 9,700 + 160 (PCARB-65)

E9 = 1,800

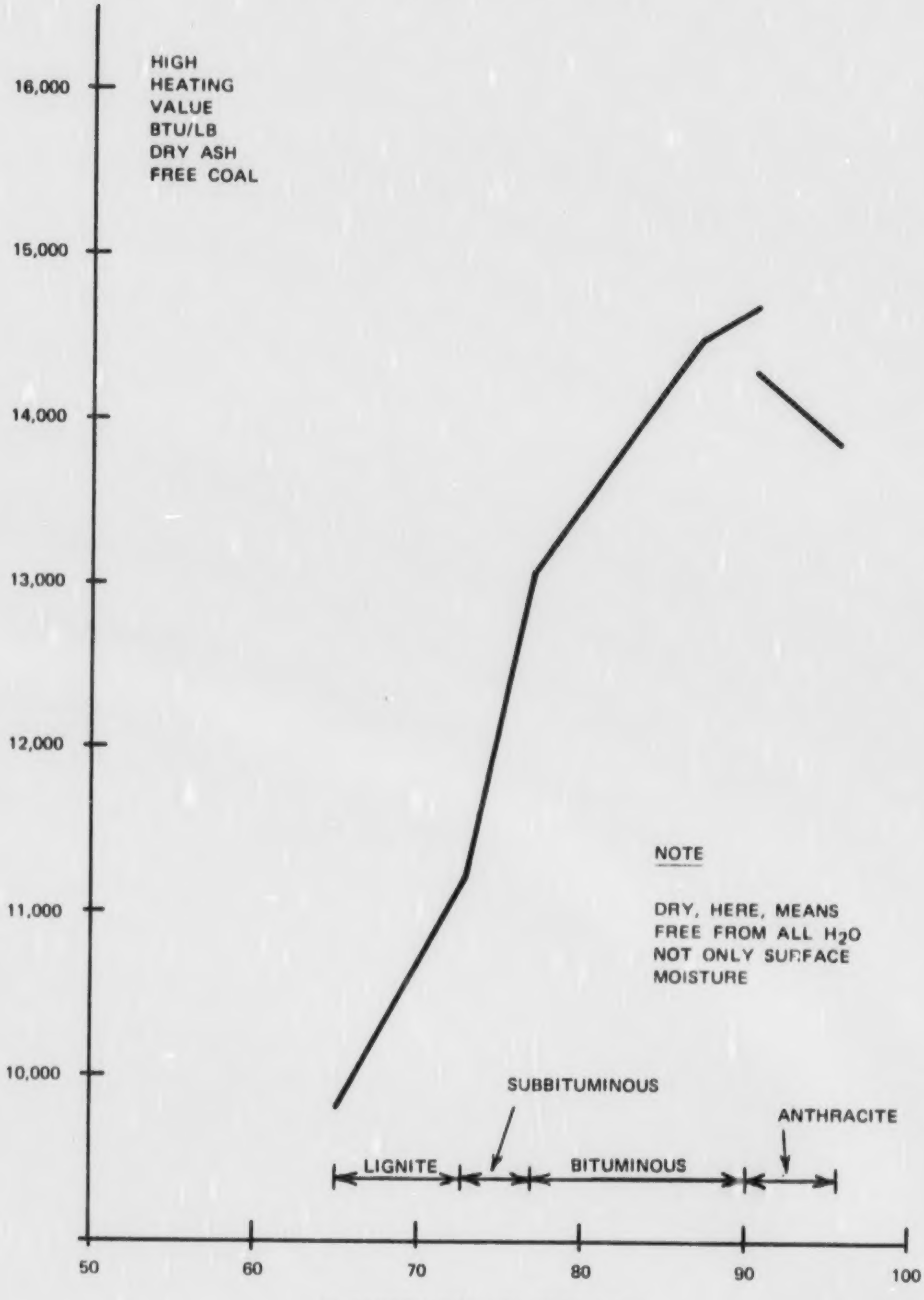
E10 = $1{,}250\ \mathrm{PSULF} + 1{,}065\ (\mathrm{TDAFC.PSULF})^{0.6} - 250$

E11 = 13,800 + 200 (PCARB-65)

E12 = 14,000

FIGURE 8.1

HIGH HEATING VALUE OF VARIOUS RANKS OF COAL



REFERENCES:
1. 1972 KEYSTONE COAL INDUSTRIAL MANUAL, MCGRAW HILL.
2. PERRY, H., "CHEMICAL ENGINEERING HANDBOOK," 4TH EDITION.
3. LOWRY, H., "CHEMISTRY OF COAL UTILIZATION," WILEY.

FIGURE 8.2

DRY ASH FREE COAL, OXYGEN AND STEAM REQUIREMENTS FOR A 250 X 10^9 BTU/DAY LURGI SNG PLANT

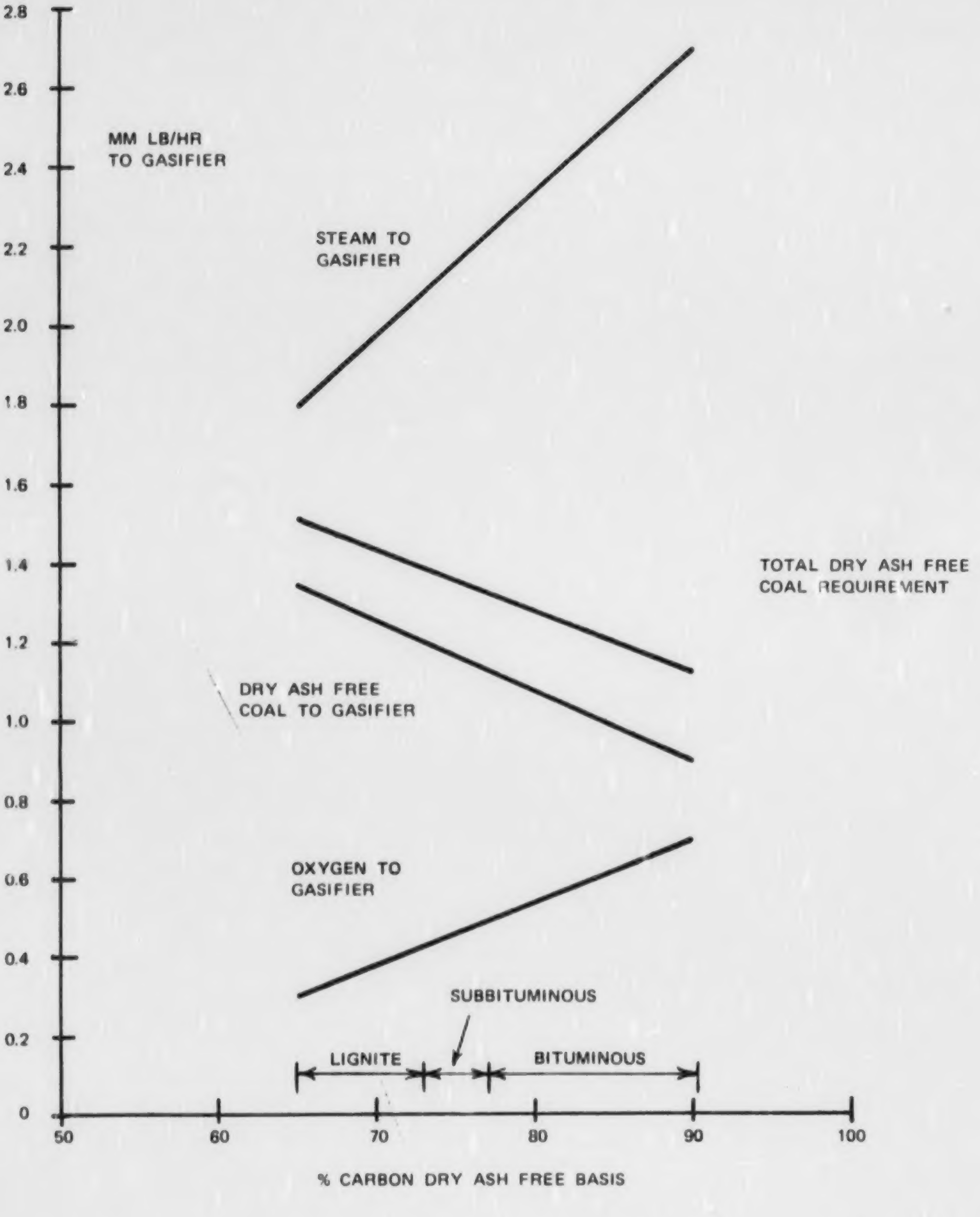


FIGURE 8.3

LURGI SNG PROCESS FLOW DIAGRAM

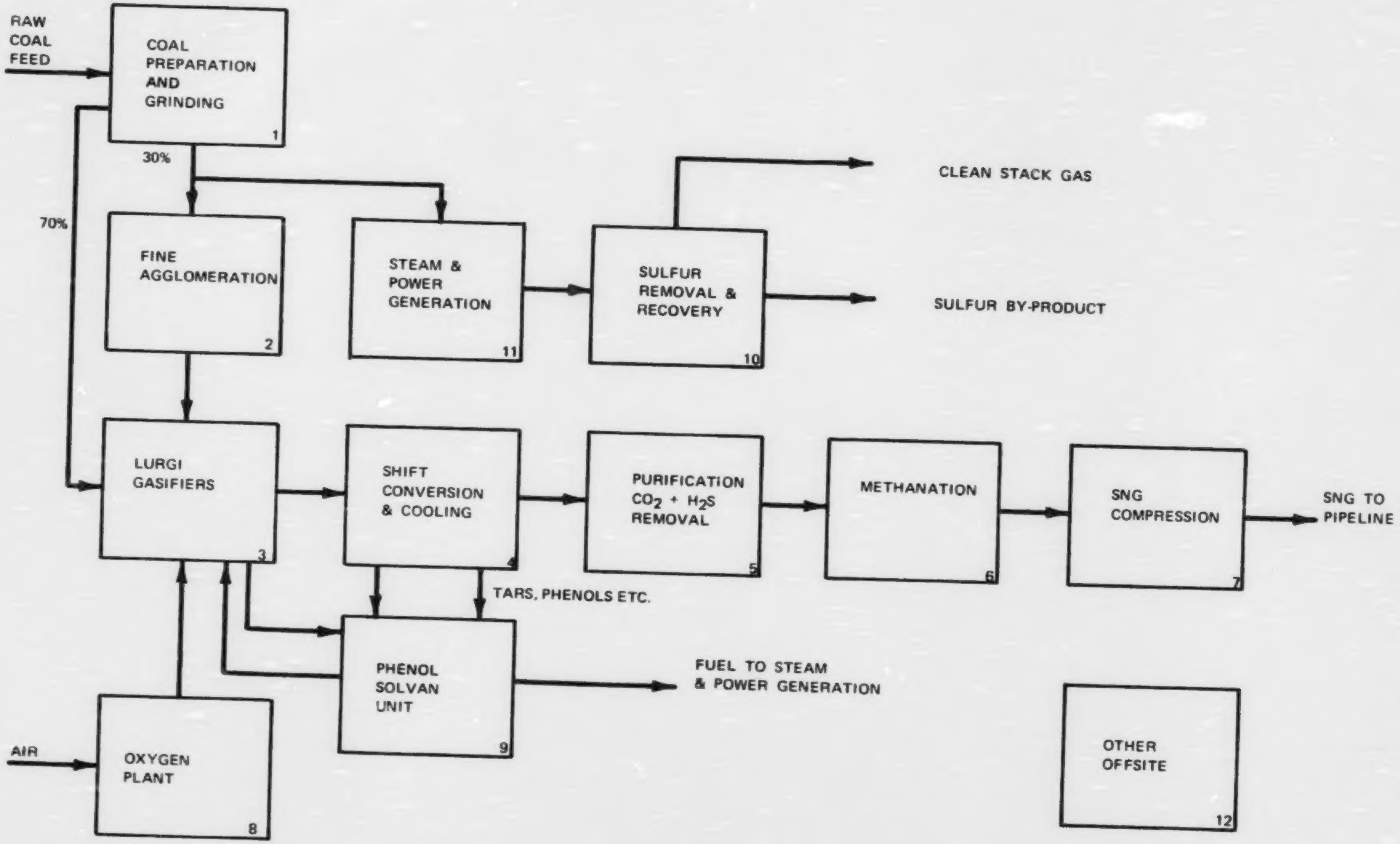

RAW COAL FEED
COAL PREPARATION AND GRINDING
1
30%
70%
FINE AGGLOMERATION
2
STEAM & POWER GENERATION
11
SULFUR REMOVAL & RECOVERY
10
CLEAN STACK GAS
SULFUR BY-PRODUCT
LURGI GASIFIERS
3
SHIFT CONVERSION & COOLING
4
PURIFICATION CO_2 + H_2S REMOVAL
5
METHANATION
6
SNG COMPRESSION
7
SNG TO PIPELINE
TARS, PHENOLS ETC.
PHENOL SOLVAN UNIT
9
FUEL TO STEAM & POWER GENERATION
AIR
OXYGEN PLANT
8
OTHER OFFSITE
12

FIGURE 8.4

EFFECT OF LOCATION FACTOR ON GAS COST

SUBBITUMINOUS COAL % CARBON 74 DAFB
% SULFUR 0.1 DAFB

MINE-MOUTH COAL COST 6 $/TON
4.5 $/TON
3 $/TON

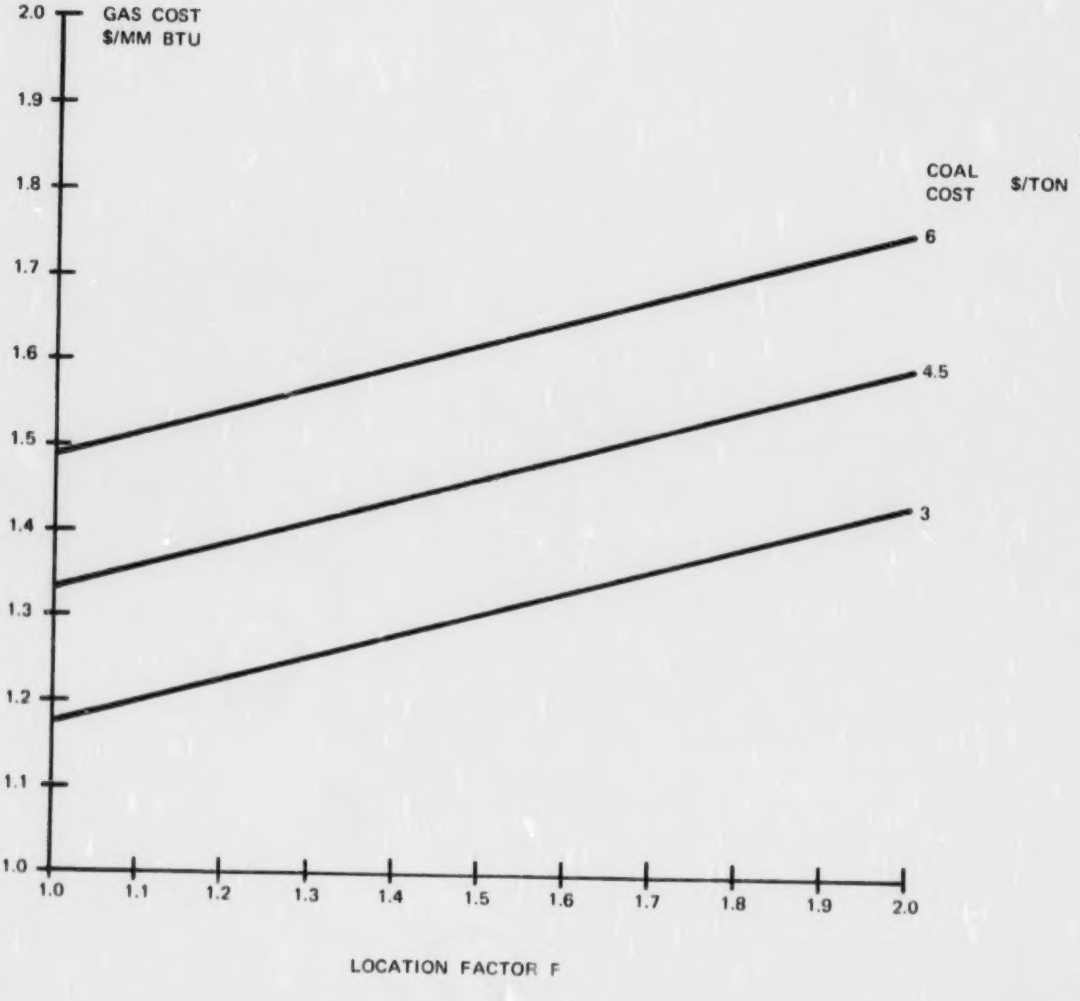


FIGURE 8.5

EFFECT OF LOCATION FACTOR ON GAS COST

BITUMINOUS COAL % CARBON 78 DAFB
% SULFUR 6 DAFB
2 DAFB

MINE-MOUTH COAL COST $6/TON
$3/TON

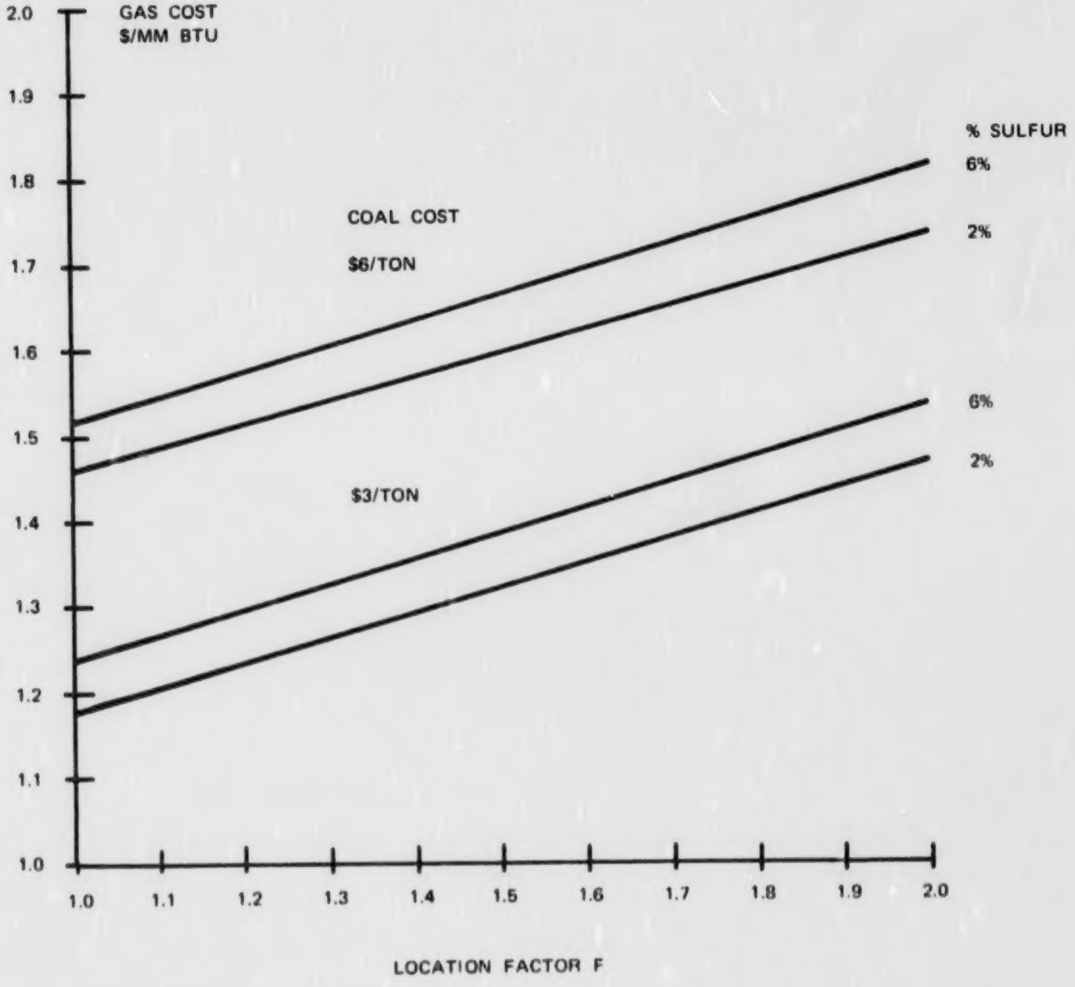


FIGURE 8.6

EFFECT OF CARBON CONTENT OF COAL ON SNG COST

% SULFUR 0, 1, 2, 4, 6

MINE-MOUTH COAL COST $6/TON
$3/TON

LOCATION FACTOR F = 1.5

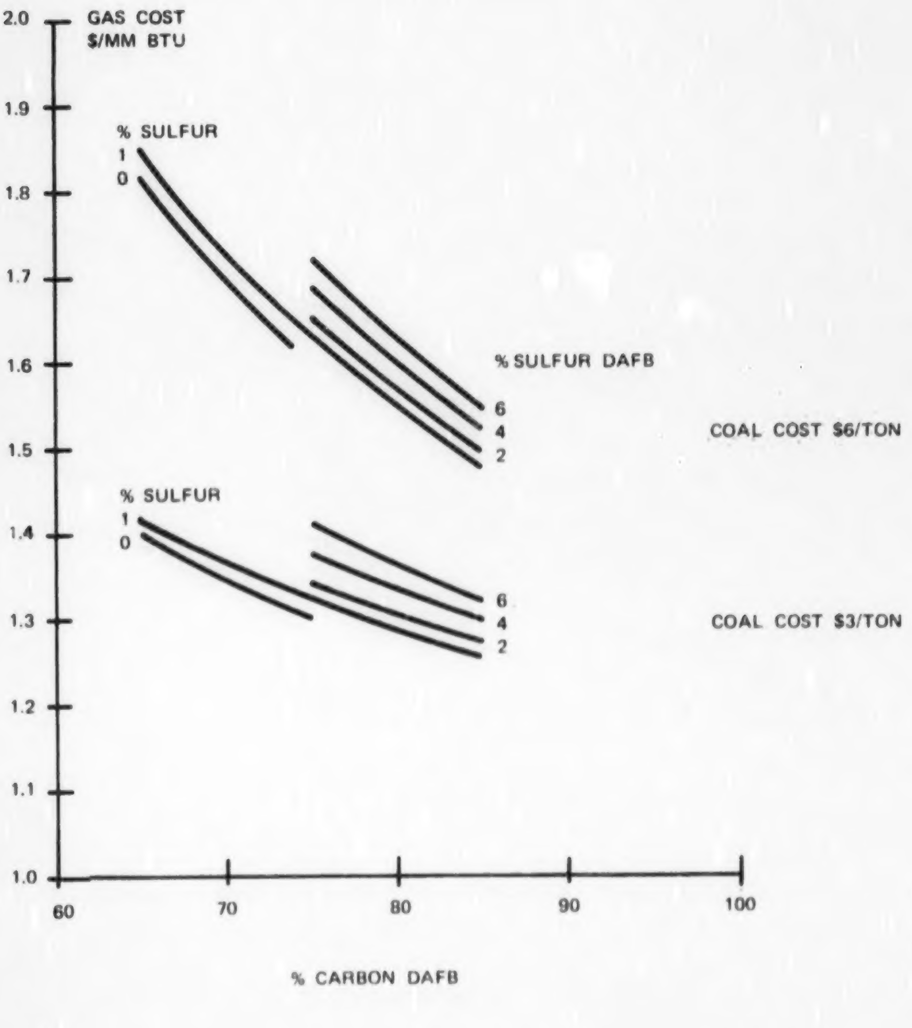
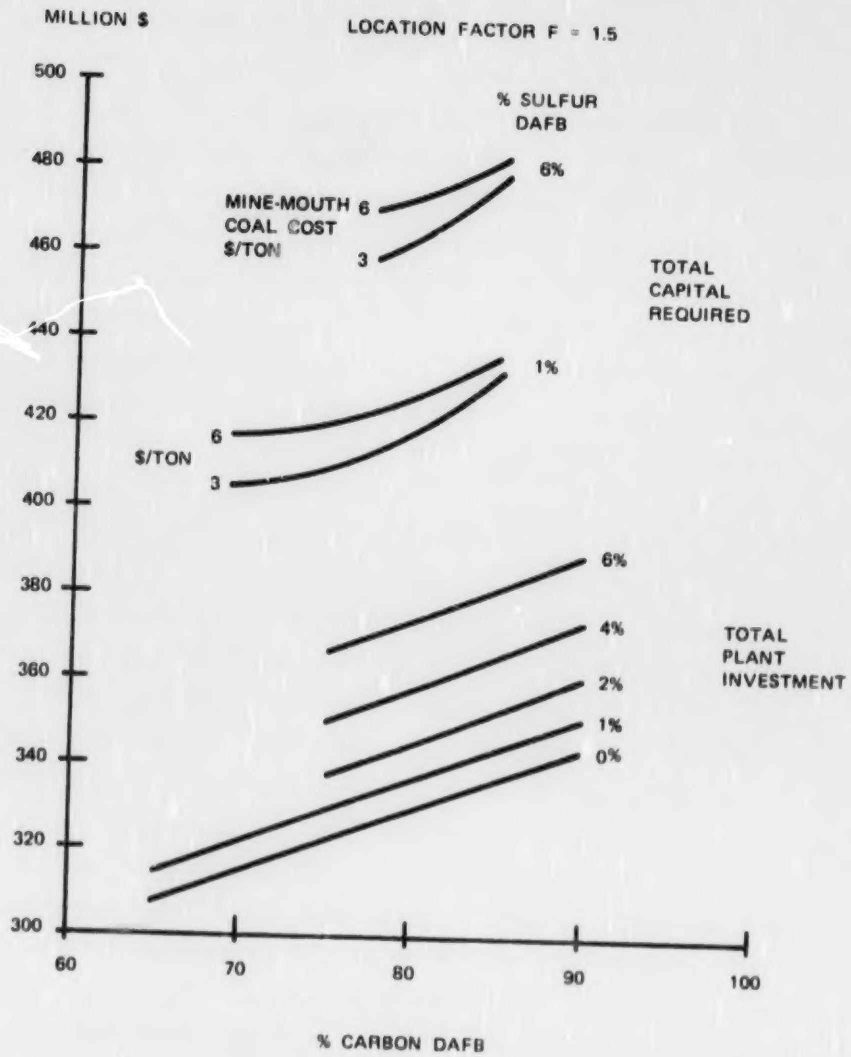


FIGURE 8.7

EFFECT OF CARBON CONTENT OF COAL ON SNG CAPITAL COSTS



9. SOLVENT REFINED COAL PRODUCTION

9.1 Process Appraisal

The Stearns-Roger Corporation's Report of July 1969, prepared for the Pittsburg and Midway Coal Mining Company was chosen as the basis for the development of this model (30, 32). Their design was examined by MWK Research and Engineering Development Department and found to be unrealistic in some areas. A number of process modifications have been made and these are discussed below.

Their design imports a quantity of natural gas equivalent to the heat content of 10% of the total solvent refined coal product. This gas is used for the hydrogen plant feed and for fuel. Under present conditions this is basically unsound. Although the price of natural gas is still low at present, it is more flexible and has a lower sulfur content than the solvent refined coal and is potentially a more valuable product. The process has, therefore, been modified to use the light oil and hydrocarbon by-product streams as fuel and hydrogen plant feed.

In the Stearns-Roger Design, a fluidized combustor was included to burn off the carbon remaining on the ash. An examination was made by MWK to determine the economics of burning off this carbon. The total plant investment of the carbon burn-off section was $9 million and the total annual production cost of the electricity produced by the plant was $2.1 million. The electricity could only be sold for $1.3 million, so this area of the plant would operate at a loss. Instead, the dry ash is conveyed to storage after stripping off wash solvent. The steam, which would have been generated by burning-off the carbon, produced approximately the excess power which was to be sold. The chance of selling carbon-ash as by-product is considered slim. In fact, it is more realistic to provide an annual operating cost for the disposal of ash. The ash could be dumped as land fill where coal is mined, assuming the plant is built adjacent to the coal mine.

The process conditions were changed so that 2% more fuel was produced at the expense of producing 2% less solvent refined coal and the plant is now in energy balance. The hydrogen consumption and losses were estimated to be about 60% more than the figure given in the Stearns-Roger design and the feed to the hydrogen plant was adjusted accordingly.

The plant size was increased so that the solvent refined coal production rate is 250×10^9 Btu/day (higher heating value). This enables direct comparison to be made with the SNG process.

The revised plant flow rates are given below:

	Tons/Day
Raw Coal Feed to Plant	13,600
Solvent Refined Coal Product	7,834
(Equivalent HHV of SRC = 250 billion Btu/day)	
Sulfur by-product (LT/D)	300
Cresylic acid by-product	170

The revised heat requirements and fuel production are given below:

Consumption	MMBtu/hr
Dissolver preheaters	1,770
Vacuum flash preheater	450
Wash solvent splitter heater	410
Ash residue drying	70
Power generation	50
Hydrogen plant fuel	680
Hydrogen plant feed	1,150
Miscellaneous	70
	4,650
Production	
Fuel Gas	3,260
Light Oil burned as fuel	1,390
	4,650

If the coal feed has about 4% sulfur DAFB, then the solvent refined coal product would usually contain not less than 1% sulfur. (According to information from EPA, the sulfur content of solvent refined coal could be reduced to as low as 0.4% with process modification at additional cost.) This particular design produces a solid refined coal, which can be pulverized or sold as briquettes, or a liquid product depending on the need of the plant.

9.2 Process Description

The process model for the solvent refined coal process has been developed as 9 sections (Fig. 9.2). A brief description is therefore given for each of these sections.

Section 1

Raw coal from storage is first crushed and then processed through a secondary grinder to reduce the coal particles to less than 1/8 of an inch. The resulting coal fines pass through a flash dryer to remove the moisture content.

Section 2

The coal together with the solvent and hydrogen are passed through preheaters and dissolvers. The coal dissolves in the solvent in the presence of hydrogen at 1000 psig and 825°F. The dissolution of coal involves hydrogenation and depolymerization. The remaining undissolved material consists of the ash content of the coal. This section also includes the hydrogen compressors.

Section 3

The ash residue from the coal is separated from the solvent by rotary filters at 150 psig and 600°F. The ash portion is transfered to the ash drying section for further solvent recovery and on to storage.

Section 4

The solvent and light oils are recovered by a series of flash separators followed by vacuum distillation. The overheads are further distilled to recycle the solvent and produce a light fuel oil. The vacuum tower bottom product is the liquid refined coal. This section also includes a cresylic acid recovery unit.

Section 5

In this section the liquid coal product is solidified and transfered to storage.

Section 6

This section generates makeup hydrogen for the process by steam reforming the light oil stream.

Section 7

The fuel gas and light fuel oil are treated to remove hydrogen sulfide and sulfur compounds. The hydrogen sulfide goes to the sulfur recovery unit, which produces a saleable by-product.

Section 8

The steam and power generation plant is fired by fuel gas and light fuel oil.

Section 9

This section includes other offsites: the cooling water system, water treatment and general plant buildings.

9.3 Cost Model

9.3.1 Total Plant Investment

The costs presented in the Stearns-Roger Report (30) were updated to the end of 1973 and adjusted using the factors listed in the General Cost Model to represent the increased plant size. For this model no estimating has been carried out by MWK and the costs have therefore only been presented as plant investments for each area. The cost of each section has been examined and approximate adjustments have been made to the estimates in those areas whose cost appeared to be low. There is general agreement with the Stearns-Roger's cost in the following areas: raw coal preparation, ash filtration and drying, product solidification, hydrogen plant, fuel treatment and sulfur recovery, and steam/electricity generation. However, it was felt that the other areas, viz., the preheater/dissolver units, solvent/light oil/cresylic acid recovery and the general offsite units were on the low side.

The preheater/dissolver units are large items operating at high pressures and temperatures. Much of the material of construction is stainless steel. The dissolver design is complex and not very well defined as yet. The plant investment of this area has therefore been increased over the adjusted Stearns-Roger's cost by $10 million.

The solvent/light oil/cresylic acid recovery units are relatively complex, large units largely constructed of stainless steel. The plant investment for this area has been increased by $15 million over the Stearns-Roger figure.

It was felt that $30 million for other offsite units was more likely than the adjusted Stearns-Roger Figure of under $10 million.

In making these adjustments to the Stearns-Roger costs, it should be stressed that no detailed estimating was done by MWK and these changes should be regarded as approximate only.

The Total Plant Investment for a plant producing 250×10^9 Btu/day of solvent refined coal is given below.

Section Number	Section Description	Total Plant Investment (TPI) F=1.0 M$
1	Coal preparation (solid handling section)	10,000
2	Preheater/dissolvers	40,000
3	Ash filtration, drying and disposal	15,000
4	Solvent/light oil/cresylic acid recovery	30,000
5	Product solidification/handling and storage	10,000
6	Hydrogen plant	10,000
7	Sulfur removal from fuels and sulfur recovery	10,000
8	Steam and power generation	10,000
9	Other offsites	30,000
		165,000

If F = 2.0, the value of TPI is $215 million.

9.3.2 Total Net Annual Operating Cost, Total Capital Requirement and Total Annual Production Cost

The total net operating cost, AOC, is the total cost of operating the plant less the credits from the sale of by-products. It does not include return of capital, payment of interest on debt or income tax on equity returns and is given by:

$$\underline{\underline{AOC = 0.078\ TPI + 2TO.CO(1+F) + ANR}}$$

The total number of shift operators for a plant producing 250×10^9 Btu/day of solvent refined coal is estimated to be 200.

The annual cost of raw materials less by-product credits (ANR) is given by:

$$\underline{\underline{ANR = ACOAL + ACHEM - ASULF - ACRES}}$$

The annual cost of coal feed to the plant (ACOAL) is given by:

$$\underline{\underline{ACOAL = CCOAL \cdot TCOAL \cdot SD}} \quad M\$/Year$$

where CCOAL is the cost of coal in \$/ton, TCOAL is the total raw coal feed to the plant in Mton/day and SD is the number of days the plant is on stream per year.

The annual cost of catalyst and chemicals, ACHEM, can be assumed to be 500 M\$ for the plant considered. The annual credit for the sale of sulfur, ASULF, is given by:

$$\underline{\underline{ASULF = CSULF \cdot TSULF \cdot SD}} \quad M\$/Year$$

where CSULF is the unit credit for sulfur in \$/LT and TSULF is the sulfur production rate in MLT/day.

The annual credit for the sale of cresylic acid is given by:

$$\underline{\underline{ACRES = CCRES \cdot TCRES \cdot SD}} \quad M\$/Year$$

where CCRES is the unit credit for cresylic acid and TCRES is the production rate of cresylic acid in Mton/day.

For the present plant using a 4% sulfur coal (DAF), TSULF is estimated to be 300 LT/day and TCRES, 170 tons/day.

The Total Capital Required, TCR, including interest on construction capital, startup costs and working capital is given in the General Cost Model as:

$$\mathrm{TCR} = 1.21\ \mathrm{TPI} + 0.8\ \mathrm{TO}\cdot\mathrm{CO}\ (1+F) + 0.4\ \mathrm{ANR}$$

The Total Annual Production Cost, TAC, including the return of capital, payment of interest and income tax on equity return is given by:

$$\mathrm{TAC} = 0.225\ \mathrm{TPI} + 2.1\ \mathrm{TO}\cdot\mathrm{CO}\ (1+F) + 1.04\ \mathrm{ANR}$$

9.3.3 Calculation of Costs of Solvent Refined Coal

Location Factor: F = 2.0

Coal Details: Bituminous PCARB = 78%, PSULF = 3.8%

Plant Details: 250 billion Btu/day of SRC
Coal Feed rate 13,600 tons/day
On Stream for 340 days/year
Total number of shift operators TO = 200

By-Products: 300 tons/day of sulfur at $5/ton
170 tons/day of cresylic acid at $100/ton

Example 1

Coal Cost: $3/ton

ACOAL = 13.600 x 240 x 3 = M$ 13,900
ASULF = .300 x 340 x 5 = M$ 500
ACRES = .170 x 340 x 100 = M$ 5,700
ACHEM = Cost of catalysts and chemicals = M$500

The cost of raw materials and chemicals less by-product credits is given by:

ANR = 13,900 + 500 - 500 - 5,700

ANR = M$8,200

TAC = 0.225 x 215,000 + 2.1 x 200 x 7 (1+2) + 1.04 x 8200

TAC = $65.7 million/Yr

$$\text{Cost of SRC} = \$65.7 \times 10^{6}/(340 \times 250{,}000\ \text{MMBTU}) = \$0.77/\text{MMBtu}$$

Example 2

Coal Cost: $6/ton

ACOAL = M$27,800

ANR = 27,800 + 500 - 500 - 5,700

ANR = M$22,100

TAC = 0.225 x 215,000 + 2.1 x 200 x 7 (1+2) + 1.04 x 22,100

TAC = $80.2 million/Yr

$$\text{Cost of SRC} = \$80.2 \times 10^{6}/(340 \times 250{,}000\ \text{MMBtu}) = \$0.94/\text{MMBtu}$$

Figure 9.1 illustrates the variation in the cost of solvent refined coal with location factor, for coal costs of $3 and $6/ton. The costs of substitute natural gas produced by a plant of the same size using the same bituminous coal feedstock are also given for comparison.

An underestimation of 30% in the total plant investment would result in an underestimation of SRC costs by 18 to 24¢/MMBtu depending on the location. Because of the limited state of development of the SRC process, the order of accuracy of the TPI estimate is only about 30%.

9.4 Conclusions

Solvent refined coal can be produced more cheaply than SNG. It is, however, a solid fuel containing normally about 1% sulfur when produced from a 4% sulfur coal (DAF). According to proprietary information from EPA, SRC with as low as 0.4% sulfur can be produced with process modification and increase in cost. Presumably this can be done by increasing the hydrogenation pressure in the dissolvers. However the main area that needs to be improved seems to be the ash filtering section. This section is probably the most costly as well as troublesome in operation.

Another aspect of investigation for the production of SRC appears to be the market. It is basically an expensive, low sulfur, ash free fuel not suitable for direct use with gas turbines. Since the process can also be geared towards specialized refinery type products, the question arises as to whether this would be a more worthwhile direction than the production of a solid fuel.

FIGURE 9.1

COMPARISON OF SOLVENT REFINED COAL AND SUBSTITUTE NATURAL GAS COSTS BITUMINOUS COAL FEED, 3.8% SULFUR

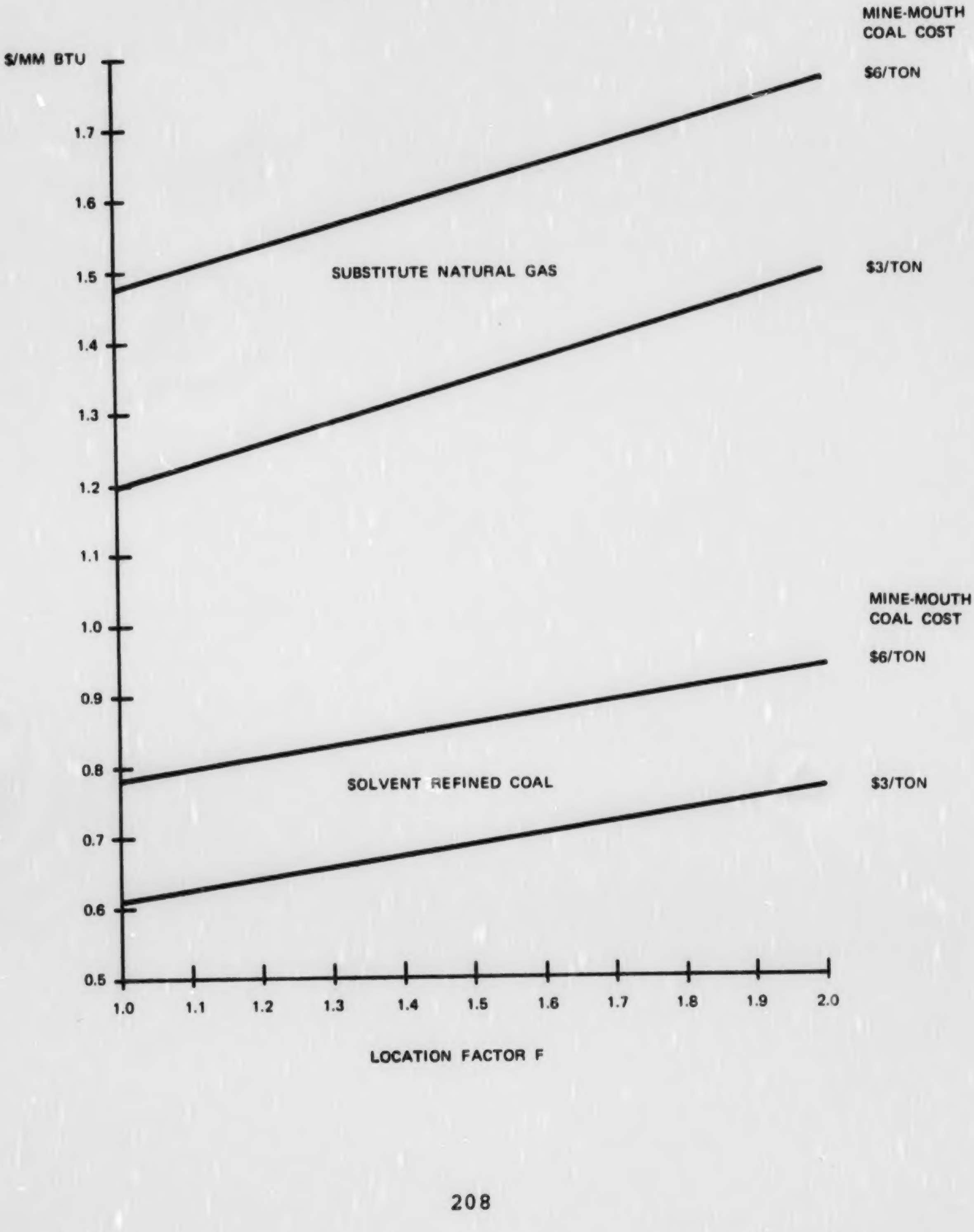
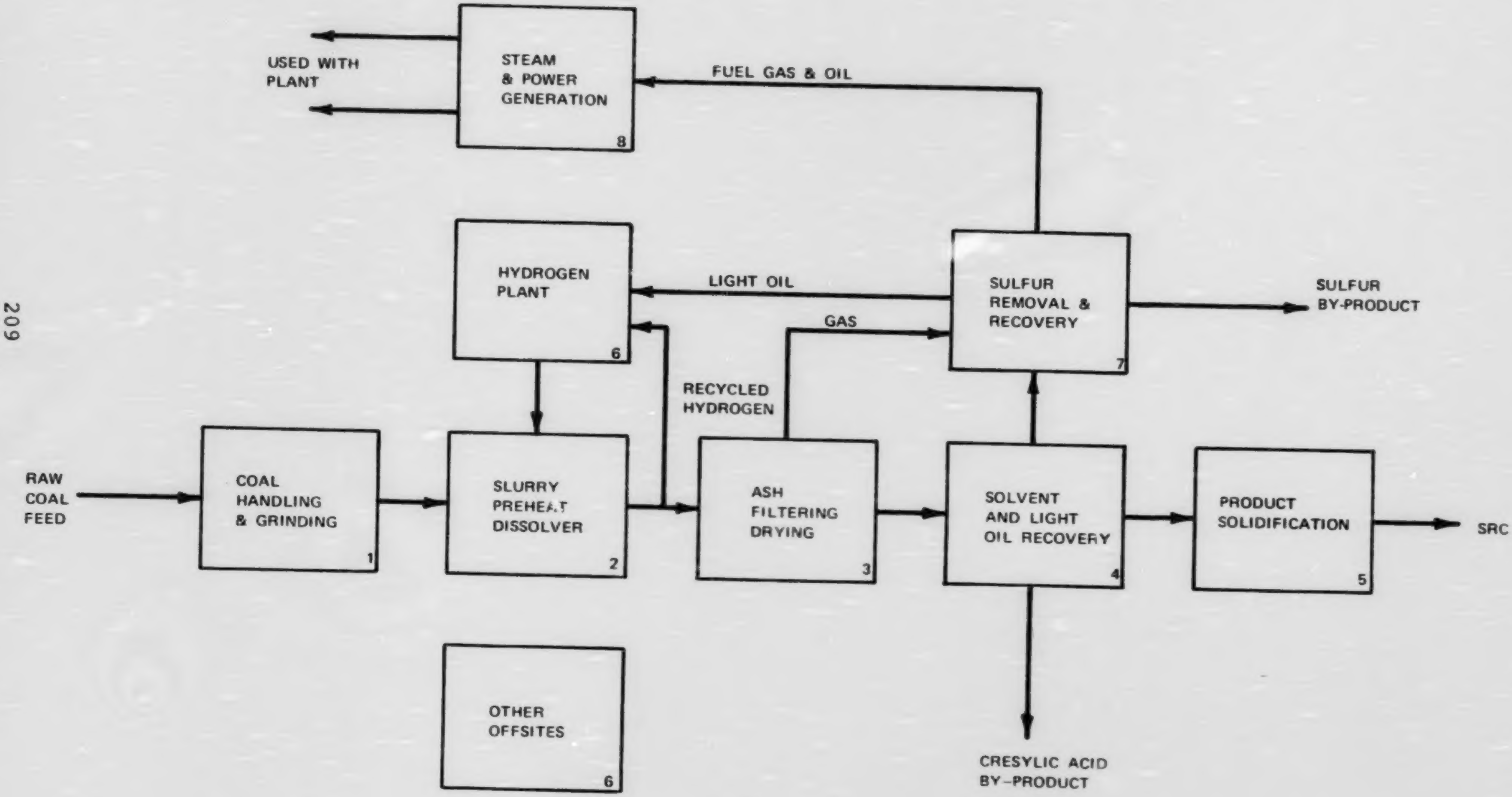


FIGURE 9.2

SOLVENT REFINED COAL PROCESS FLOW DIAGRAM


USED WITH PLANT
STEAM & POWER GENERATION
8
FUEL GAS & OIL
HYDROGEN PLANT
6
LIGHT OIL
GAS
SULFUR REMOVAL & RECOVERY
7
SULFUR BY-PRODUCT
RECYCLED HYDROGEN
RAW COAL FEED
COAL HANDLING & GRINDING
1
SLURRY PREHEAT DISSOLVER
2
ASH FILTERING DRYING
3
SOLVENT AND LIGHT OIL RECOVERY
4
PRODUCT SOLIDIFICATION
5
SRC
OTHER OFFSITES
6
CRESYLIC ACID BY-PRODUCT

10. THE COMBINED GAS TURBINE - STEAM TURBINE POWER PLANT USING A LOW BTU LURGI GASIFIER

10.1 Introduction

A new coal fired conventional steam cycle power station without stack gas cleaning can achieve a heat rate as low as 8,600 Btu/ net kwh which is equivalent to an overall cycle efficiency (based on the higher heating value of the coal) of almost 40%. If this power station was designed with, say, a Wellman/Allied stack gas scrubbing system and was burning coal with 4% sulfur DAFB, the fuel required would be increased by more than 11%, using the present design. The heat rate would be increased to more than 9,600 Btu/ net kwh and the overall cycle efficiency (net power/coal input) reduced to 35.5%. If sulfur dioxide emission controls are to be imposed on new power stations there is clearly a great deal of incentive to investigate alternatives to the conventional steam cycle power plant with stack gas scrubbing. One such alternative is a coal gasification plant supplying clean, hot, low Btu fuel gas under pressure to a combined gas turbine - steam turbine power generation unit.

There are many coal gasification processes in various degrees of development. The Lurgi coal gasification unit using either air or oxygen is, however, well established on a commercial scale. For this reason, in these studies, the Lurgi coal gasification unit, using air, has been selected to provide the low Btu fuel gas. Emphasis has been given to designs which are possible at present, although calculations have been made for future cycles not limited by the gas turbine inlet temperature.

When this work was started it was felt that a gas turbine inlet temperature of 1700°F was the highest allowable design temperature for base loaded plants. Recent discussions (December 1973) with General

Electric Company (39) reveal however, that they now have marketed a base load air compressor/gas turbine/generator unit producing 55 net megawatts, operating with an inlet temperature of 1950°F and an inlet pressure of 150 psia. To some extent this makes cycles 1 and 2 (to be defined later) already outdated; however, the results and discussion on them are included for general interest.

These power plants are obviously base loaded, since the Lurgi gasifiers cannot be shut down other than for maintenance. The gas turbine must therefore be capable of in excess of 8000 hours operation per year. It is also absolutely necessary that the inlet gas and air have practically zero alkali metal impurities. A special water wash, free of sodium and potassium ions is therefore needed after the hot potassium carbonate purification plant.

A gas turbine inlet temperature of 1600°F obviously puts some limitations on the design of the power cycle, since the exhaust temperature is less than 1000°F (gas turbine exhausts at 16 psia). The most widely used power plant steam cycle has 1000°F, 2400 psia steam with reheat to 1000°F after expansion down to 500-600 psia (33, 38). This steam cycle is illustrated in Figure 10.1 and is essentially the steam cycle used in all the combined cycle studies, although in these, it is clearly necessary to preheat the water with the flue gas. In a conventional power plant, this is done by bleeding steam from several pressure levels in the turbines and preheating the combustion air with the flue gas. The steam cycle illustrated in Figure 10.1 is not the most efficient possible, since steam turbine efficiencies of 88% (isentropic work to electrical power) have been used. However, this is intended to be a comparative study and in the combined power cycle work, gas turbine, steam turbine and air compressor efficiencies of 88% were used. Again it must be emphasized that the combined cycles reported in this work should not be considered as finalized designs, but taken as illustrations of the possibilities of the cycles under these stated conditions.

It is clear that with gas turbine inlet temperature of 1600°F, the steam cycle cannot receive heat only from the gas turbine exhaust, since its temperature is too low (Fig 10.16).

One method of achieving a sufficiently high temperature for the superheat and reheat tubes in the steam turbine section is to divide the fuel gas into 2 streams. One stream is combusted with excess air to give an inlet temperature of 1600°F and expanded to 16 psia and less than 1000°F. The second fuel gas stream is expanded to 16 psia and then mixed with the first gas turbine exhaust in a combustor. The combustion temperature will then be in excess of 1200°F and suitable for supplying heat to the reheat and superheat tubes. This cycle was investigated and found to produce less power than cycle 1 (illustrated in Figure 10.3). It is also more complicated and requires an extra turbine. The results for this alternative have therefore not been reported here.

Another area of work for which the results are not reported should be mentioned if only to avoid further study. There is a great temptation to assume that conventional centrifugal air compressors with intercoolers to reduce the power consumed would be used in a combined power cycle. This is not so. In fact, much higher overall cycle efficiencies are achieved by using axial flow compressors with no intercoolers. This fact has been reported in a few publications but is still not widely recognized. The use of centrifugal compressors rejects well over 1000 MMBtu/hr of heat to cooling water. A design utilizing axial flow compressors keeps the heat of compression in the power cycle.

10.2 The Lurgi Gasification Plant

Before discussing details of the power cycles, a brief description of the low Btu Lurgi gasification plant is useful. Approximately 20 Lurgi gasifiers are necessary to generate the fuel gas for a 1000 megawatt combined cycle power plant. The air enters the gasifiers at about 600°F, preheated by compression to 320 psia, and the steam enters at about 460°F from an intermediate pressure level in the steam power cycle. For the case studied a bituminous coal with the following analysis was used:

	Wt % DAFB		% As Received
C	78.0	Water	15
H	5.5	Ash	15
O	11.0		
S	4.0		
N	1.5	HHV = 14,200 Btu/lb DAF	
	100.0		

The coal feed to the gasifiers was taken as 710,000 lb/hr DAF. The significance of this number is that it is the coal feed to a new conventional steam cycle power plant fitted with a Wellman/Allied stack gas scrubbing system and generating 1000 megawatts net power. The Wellman/Allied process produces a sulfur by-product and is therefore compatible with the low Btu Lurgi design. The steam cycle is as shown in Figure 10.1 and its heat rate would be 9000 Btu/net kwh without the stack gas scrubbing unit. So in the power cycle studies a combined cycle which produces net power over 1000 megawatts is more efficient than the conventional steam cycle with Wellman/Allied stack gas scrubbing. (If the power plant were fitted with a Wet Limestone system, the net power produced would be 1074 megawatts. Combined cycles producing power in excess of this amount would be more efficient than a conventional plant with Wet Limestone scrubbing.). The conventional steam cycle without stack gas scrubbing would achieve 1127 net megawatts with this coal feed rate.

The composition of the crude gas after the tars have been removed is approximately (34):

	Mole % (Dry Basis)
CO_2	14.5
CO	15.6
H_2	21.5
CH_4	6.0
N_2	41.7
H_2S	0.7
	100.0

The gas has an additional 26% of water. It is purified by a hot carbonate scrubbing system. The gas leaves the absorber at 260°F and is reheated to 300°F by heat exchange with gas leaving the gasifier. The gas is saturated by the water wash for alkali metal removal. The reboiler heat required by the stripper tower of the hot carbonate system is supplied by cooling the tar-free gas entering the purification system from 500°F to 260°F.

The gas entering the combined power cycle has approximately the following composition:

	Mole % (Dry Basis)
CO_2	10.65
CO	16.5
H_2	22.5
CH_4	6.3
N_2	44.0
H_2S	0.05
	100.00

The gas is saturated with water at 300°F and 250 psia. The total wet flow rate is 163,300 lb moles/hour. All tars, etc. are recycled to the gasifiers after being removed in a water wash tower.

A heat balance around the Lurgi gasifiers and purification system is given in Figure 10.2.

10.3 Description of Cycles Studied

The **cycles** which were studied are illustrated in Figures 10.3-10.6. Cycle 4 (Figure 10.6) is a future cycle. Cycles 1 to 3 (Fig. 10.3-10.5) are the cycles which are possible now. The differences between them are in the positions of the preheat, superheat and reheat tubes in the steam generators. The differences between cycle 3 and cycle 4 is the gas turbine inlet pressure.

In cycle 3, gas turbine A has an inlet pressure of 100 psia and inlet temperatures greater than 1900°F. The exhaust temperatures from turbine A are therefore greater than 1200°F (see Figure 10.16) and suitable for providing heat to the whole steam cycle (see Fig. 10.14). Thus, no heat is given up to the steam cycle by the high pressure gas at the higher temperatures (the superheat & reheat tubes in Cycle 1, and the superheat tubes in Cycle 2) and the cycle is potentially more efficient. Another illustration of this principle is given by comparison of cycles 1 and 2 (Figures 10.3 and 10.4). These are different from cycle 3 in one way. The inlet temperature to gas turbine A is less than 1900°F and the exhaust temperature is too low to provide heat for the superheater and the reheat tubes of the steam cycle. Cycle 1 is not as efficient as cycle 2 (see Figure 10.7) at the same gas turbine inlet temperatures because the steam system of cycle 2 removes less heat from the high pressure hot gas. Cycle 1 has both the superheater and the reheat tubes heated by the gas before the inlet to turbine A, whereas cycle 2 only has the superheater in this location.

Examination of Figure 10.7 will also reveal another interesting point. Cycle 3 generates less power than cycle 2 when the inlet temperature to turbine A is less than 1960°F. The reason for this is that the steam cycle with 1000°F superheat and 1000°F reheat is not efficient when the gas entering the steam generator is less than 1250°F gas inlet temperature. The pinch point between the cooling curves results in the slope of the gas cooling curve being much less than that of the water preheat curve. This means that the two curves diverge and the stack gas temperature is 350°F, which is obviously undesirable, especially in view of the large amounts of excess air in the flue gas. At 1300°F gas inlet temperature the gas cooling curve is almost parallel to the water preheat curve and a stack gas temperature of 230°F is possible.

Cycles 1, 2 and 3 are identical up to the point where the gas is combusted. The clean fuel leaving the Lurgi unit at 300°F and 250 psia is preheated to 420°F by two heat exchanges with the hot compressed gasifier air. The air itself is cooled to 600°F. The fuel gas at 250 psia and 420°F is then expanded in turbine B to 105 psia and 270°F providing a power generation of 59 megawatts. The expanded fuel gas is then combusted with a quantity of air which is in excess of the stiochiometric requirements. The quantity of air which is used produces the control of the combustion temperature (see Figure 10.9). The air temperature leaving the axial flow compressor is 580°F.

Cycle 4 differs from cycle 3 in that in this case, the combustion air is compressed to 250 psia and 920°F by an axial flow compressor. Turbine B is therefore not required. Turbine A expands the combusted gas from 250 psia to 16 psia. Again cycle 4 does not produce an efficient steam cycle until the gas turbine exhaust temperature exceeds 1250°F which corresponds to an inlet temperature 2430°F.

In all four cycles the steam to the gasifier was taken from the steam cycle after the HP turbine. It was expanded down to the gasifier pressure to produce about 6 megawatts.

10.4 Discussion of Results

In view of the statements by General Electric Corporation that their present 55 net megawatt unit can operate base loaded with gas turbine inlet temperatures of 1950°F (39), it appears reasonable to claim that cycle 3 with an inlet temperature of 2000°F is the best available cycle for a plant designed in the next two years. It is possible that the best intermediate pressure between turbine A and turbine B may be higher than 100 psia. In fact the net power out of the air compressor/turbine A unit is higher for 150 psia than for 100 psia, but the higher pressure means using a steam cycle with superheat and reheat temperatures lower than 1000°F. In short the most suitable design at present appears to be close to cycle 3, but more detailed work and discussion with machine vendors would be required to produce an optimized design.

MWK's calculations show the combined cycle using a Lurgi gasification system produces about the same power as the conventional steam cycle plant without stack gas scrubbing. It is true that the steam cycle shown in Figure 10.1 is not the most efficient available, but the same is probably true of the combined cycle work. The efficiency ratio of (isentropic work to megawatts consumed or generated) has been assumed to be 88% for the air compressors, the gas turbines and the steam turbines. It is likely that the steam turbine and air compressor efficiencies could be higher than 88% and there are indications that 88% is possible for the gas turbines.

Thus, the overall Lurgi gasification plant and generation unit appears capable of efficiencies higher than 38% (net power generated divided by the higher heating value of the coal input). On occasion, Lurgi publications have expressed the overall efficiency in terms of the lower heating value of the coal. This produces a figure which is approximately 2 points higher, i.e., 38% (HHV) is equivalent to 40% (LHV).

The power (other than compressors) used by the plant itself was taken as 30 megawatts for the combined cycle and Lurgi plant, and 60 megawatts for the conventional steam cycle. The difference is due to the reduced boiler feed water pump power requirements and the fact that no induction fans are needed.

10.5 Cost Model

The design which has been used for the cost model is cycle 3 with a gas turbine inlet of 2000°F.

		Megawatts
Gas turbine power	=	1230
Air compressor power consumed	=	-605
Net generation	=	625
Steam turbine power	=	540
Total power generation	=	1165
Auxiliary power requirement	=	-30
Total net generation	=	1135

The plant will be scaled down to a 1000 net megawatt size (1030 megawatts including auxiliary power). For this size unit:

Net gas turbine/air compressor power	$= \frac{1030}{1165} \times 625 = 553$
Steam turbine power	$= \frac{1030}{1165} \times 540 = 477$
	1030
Auxiliary power requirements	-30
Total net generation	1000

Table 10.1 was prepared by examination of published Tennessee Valley Authority plant data (36, 37, 38). Information about the Bull Run Plant provided most of the data. The plant investment was updated to end of 1973 and brought to Gulf Coast cost. The size of the plant was adjusted to 1000 net megawatts. Figure 10.18 was prepared after examination of several TVA units. The cost of the power generating units was found to vary with the 0.8 power of plant size.

As shown in Table 10.1, the conventional steam cycle power station total plant investment is $200 million (Gulf Coast, end of 1973). A conventional power plant fitted with Wellman/Allied stack gas scrubbing and generating 1000 net megawatts would require an incremental boiler plant investment of $10 million and an extra $50 million for the scrubbing plant. Total plant investment for 1000 net megawatt station with Wellman/Allied stack gas scrubbing is $260 million.

Table 10.1 and the Lurgi SNG model were used to generate Table 10.2. The cost of the air compressor/gas turbine/generator unit was firmed to a certain extent by General Electric Company's approximate cost for their 55 megawatt unit, which was between $4 and $4.5 million, not installed.

The conclusion appears to be that the conventional steam cycle power plant without stack gas scrubbing would cost less than the equivalent combined cycle unit with a Lurgi gasifier, which in turn appears to cost less than the conventional station with the Wellman/Allied stack gas scrubbing system. The combined cycle power plant requires less coal feed than the conventional steam power plant fitted with stack gas scrubbing.

The costs and efficiencies shown here indicate an incentive to develop gas turbines which can handle higher inlet temperatures and pressures, thus making cycle 4 possible. The combined cycle power plant with a Lurgi gasification unit merits a more detailed technical and cost examination.

TABLE 10.1

Cost of a 1000 Net Megawatt Conventional Power Station

	Total Plant Investment, TPI (M$)	% of TPI
Boiler plant equipment	102,000	51
Turbo-generator unit	46,000	23
Land and structures	26,000	13
Accessory electrical equipment	14,000	7
Transmission plant	8,000	4
Miscellaneous equipment	4,000	2
	200,000	100

Notes:

1) Coal-fired plant
2) Costs are Gulf Coast, end of 1973
3) No stack gas scrubbing

TABLE 10.2

COST OF A 1000 NET MEGAWATT COMBINED CYCLE POWER PLANT

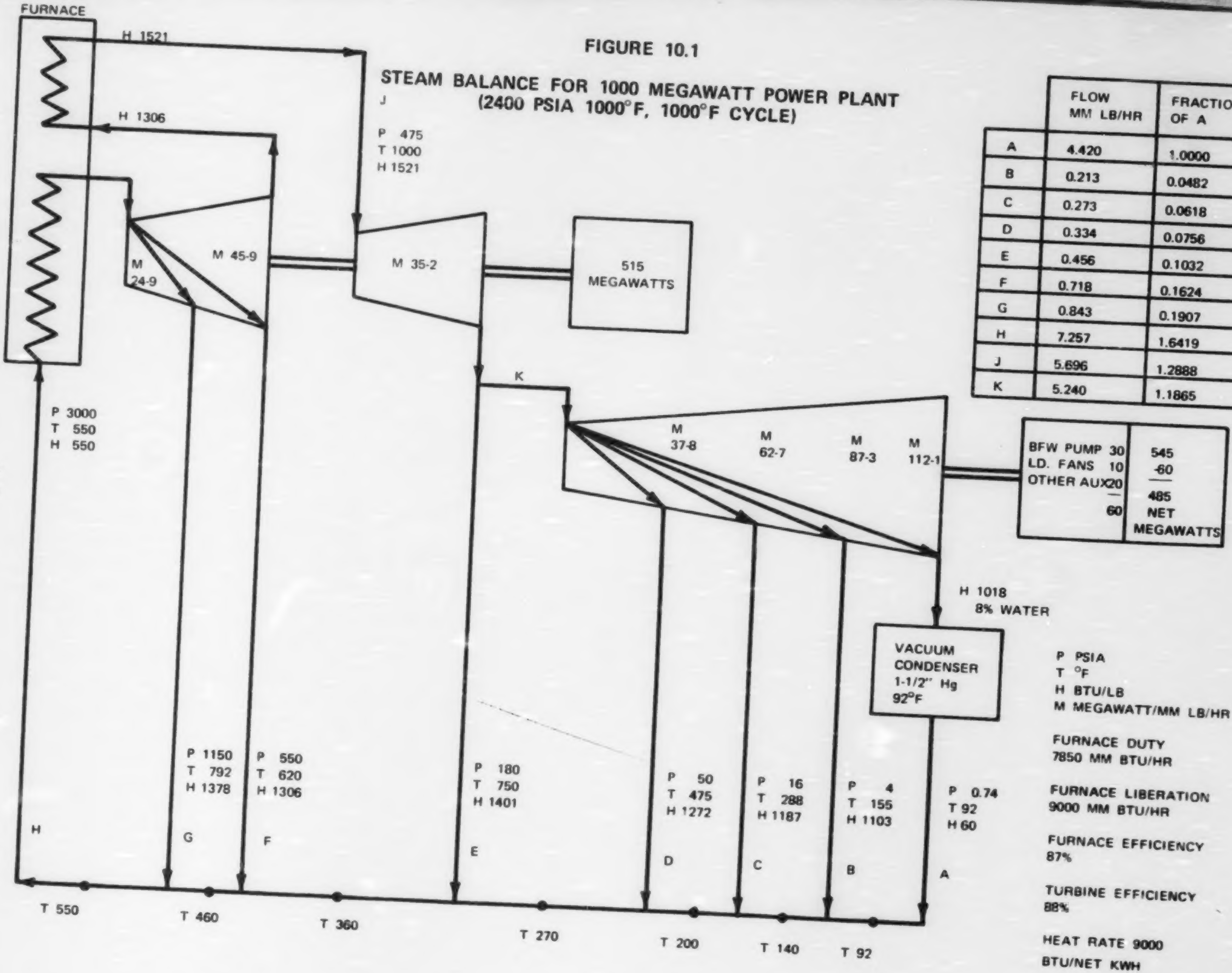
Power Generation	TOTAL PLANT INVESTMENT, TPI (M$)
Boiler plant including extra capacity for Lurgi gasifier steam. (No coal handling, ash handling or draft equipment needed)	38,000
Steam-turbo generators	25,000
Air compressor/gas turbine/generator	50,000
Land and structures	26,000
Accessory electrical equipment	14,000
Transmission plant	8,000
Miscellaneous equipment	4,000
	165,000
Lurgi Plant	
Coal preparation and handling	3,500
Fines agglomeration	8,500
Gasification	23,500
Gas Purification	6,500
Sulfur control	7,000
Offsites	16,000
	65,000

Total for gasification unit and power station = $230 million

NOTE: Costs are Gulf Coast, end of 1973

FIGURE 10.1

STEAM BALANCE FOR 1000 MEGAWATT POWER PLANT (2400 PSIA 1000°F, 1000°F CYCLE)


FURNACE
H 1521
H 1306
J
P 475
T 1000
H 1521
M 24-9
M 45-9
M 35-2
515 MEGAWATTS
K
M 37-8
M 62-7
M 87-3
M 112-1
BFW PUMP 30
LD. FANS 10
OTHER AUX20
60
545
-60
485
NET MEGAWATTS
H 1018
8% WATER
VACUUM CONDENSER
1-1/2" Hg
92°F
P 3000
T 550
H 550
P 1150
T 792
H 1378
P 550
T 620
H 1306
P 180
T 750
H 1401
P 50
T 475
H 1272
P 16
T 288
H 1187
P 4
T 155
H 1103
P 0.74
T 92
H 60
H
G
F
E
D
C
B
A
T 550
T 460
T 360
T 270
T 200
T 140
T 92

	FLOW MM LB/HR	FRACTION OF A
A	4.420	1.0000
B	0.213	0.0482
C	0.273	0.0618
D	0.334	0.0756
E	0.456	0.1032
F	0.718	0.1624
G	0.843	0.1907
H	7.257	1.6419
J	5.696	1.2888
K	5.240	1.1865

P PSIA
T °F
H BTU/LB
M MEGAWATT/MM LB/HR

FURNACE DUTY
7850 MM BTU/HR

FURNACE LIBERATION
9000 MM BTU/HR

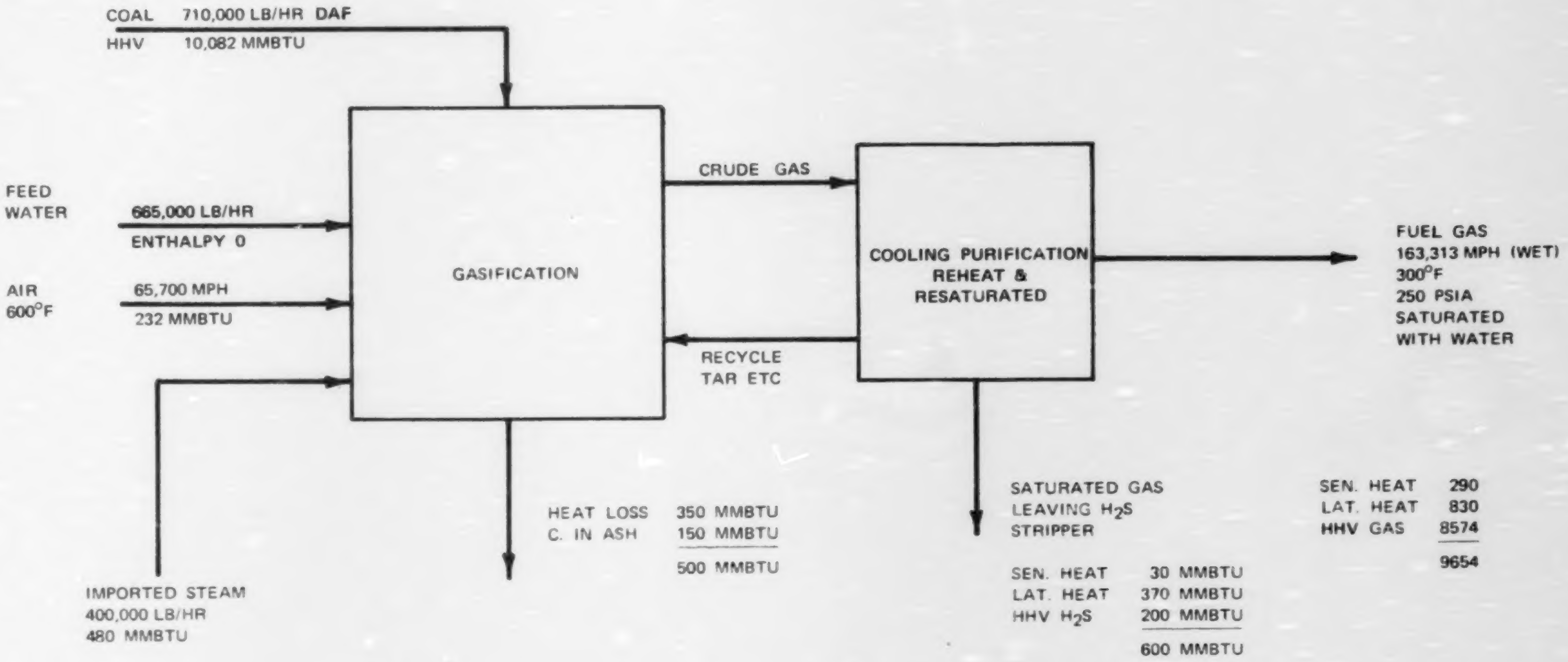
FURNACE EFFICIENCY
87%

TURBINE EFFICIENCY
88%

HEAT RATE 9000
BTU/NET KWH

FIGURE 10.2

HEAT BALANCE AROUND LURGI LOW BTU GASIFIER

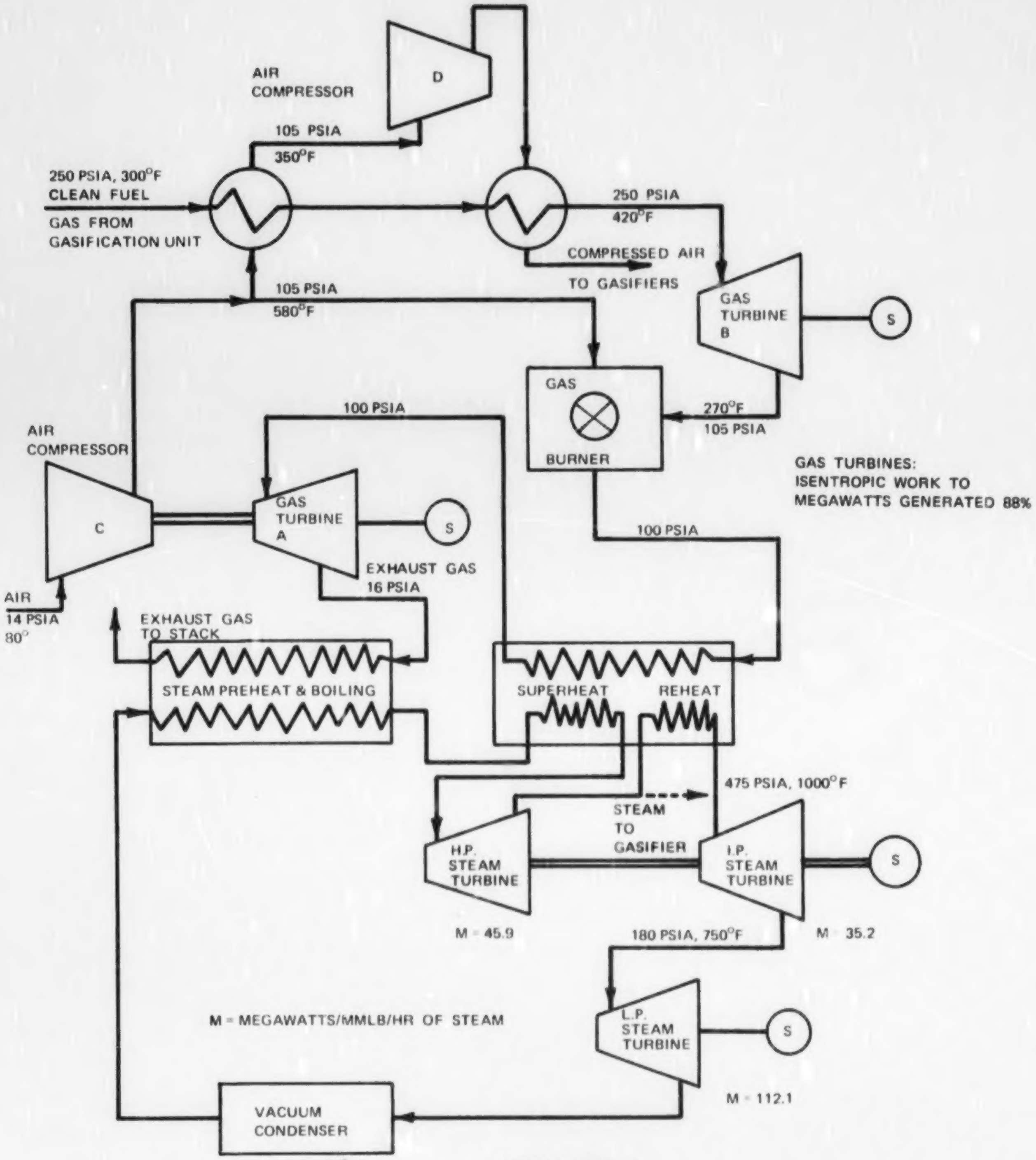

COAL 710,000 LB/HR DAF
HHV 10,082 MMBTU
FEED WATER
665,000 LB/HR
ENTHALPY 0
AIR 600°F
65,700 MPH
232 MMBTU
IMPORTED STEAM
400,000 LB/HR
480 MMBTU
GASIFICATION
CRUDE GAS
RECYCLE TAR ETC
COOLING PURIFICATION REHEAT & RESATURATED
FUEL GAS
163,313 MPH (WET)
300°F
250 PSIA
SATURATED WITH WATER
HEAT LOSS 350 MMBTU
C. IN ASH 150 MMBTU
500 MMBTU
SATURATED GAS LEAVING H2S STRIPPER
SEN. HEAT 30 MMBTU
LAT. HEAT 370 MMBTU
HHV H2S 200 MMBTU
600 MMBTU
SEN. HEAT 290
LAT. HEAT 830
HHV GAS 8574
9654

	IN
COAL	10,082 MMBTU
STEAM	480 MMBTU
AIR	232 MMBTU
	10,794 MMBTU

	OUT
PURIFIED GAS	9,694 MMBTU
STRIPPED O/H	600 MMBTU
HEAT & C. LOSS	500 MMBTU
	10,794 MMBTU

FIGURE 10.3

ADVANCED POWER CYCLE
POWER GENERATION SECTION (CYCLE 1)


AIR
COMPRESSOR
D
105 PSIA
350°F
250 PSIA, 300°F
CLEAN FUEL
GAS FROM
GASIFICATION UNIT
250 PSIA
420°F
COMPRESSED AIR
TO GASIFIERS
105 PSIA
580°F
GAS
TURBINE
B
S
GAS
BURNER
270°F
105 PSIA
100 PSIA
AIR
COMPRESSOR
C
GAS
TURBINE
A
S
EXHAUST GAS
16 PSIA
100 PSIA
AIR
14 PSIA
80°
EXHAUST GAS
TO STACK
STEAM PREHEAT & BOILING
SUPERHEAT
REHEAT
475 PSIA, 1000°F
STEAM
TO
GASIFIER
H.P.
STEAM
TURBINE
I.P.
STEAM
TURBINE
S
M = 45.9
M = 35.2
180 PSIA, 750°F
L.P.
STEAM
TURBINE
S
M = 112.1
VACUUM
CONDENSER
1-1/2"Hg. 92°F

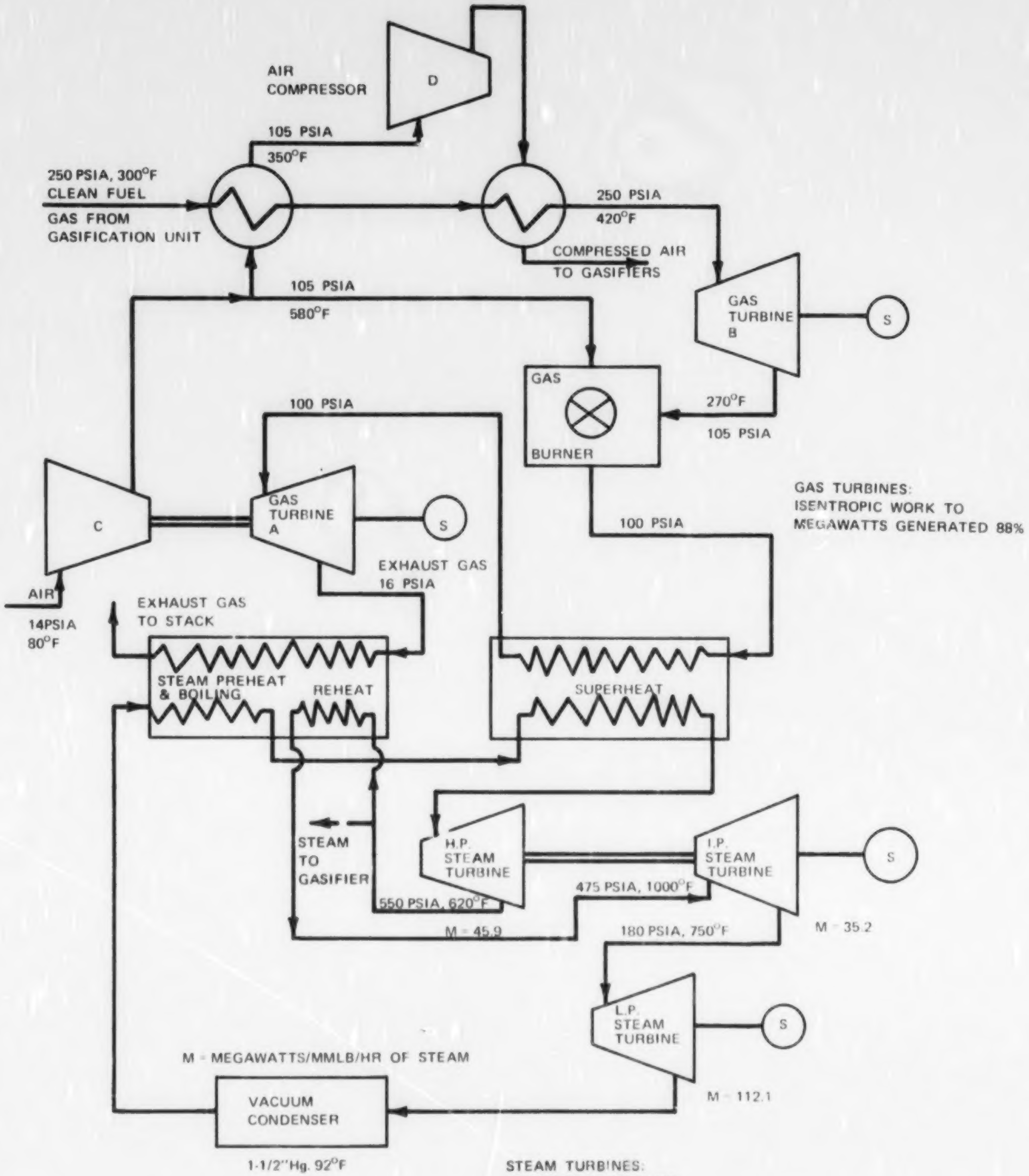
GAS TURBINES:
ISENTROPIC WORK TO
MEGAWATTS GENERATED 88%

M = MEGAWATTS/MMLB/HR OF STEAM

STEAM TURBINES:
ISENTROPIC WORK TO
MEGAWATTS GENERATED 88%

FIGURE 10.4

ADVANCED POWER CYCLE
POWER GENERATION SECTION (CYCLE 2)


AIR
COMPRESSOR
D
105 PSIA
350°F
250 PSIA, 300°F
CLEAN FUEL
GAS FROM
GASIFICATION UNIT
250 PSIA
420°F
COMPRESSED AIR
TO GASIFIERS
105 PSIA
580°F
GAS
TURBINE
B
S
GAS
BURNER
270°F
105 PSIA
100 PSIA
C
GAS
TURBINE
A
S
100 PSIA
AIR
14PSIA
80°F
EXHAUST GAS
16 PSIA
EXHAUST GAS
TO STACK
STEAM PREHEAT
& BOILING
REHEAT
SUPERHEAT
STEAM
TO
GASIFIER
H.P.
STEAM
TURBINE
I.P.
STEAM
TURBINE
S
550 PSIA, 620°F
475 PSIA, 1000°F
M = 45.9
180 PSIA, 750°F
M = 35.2
L.P.
STEAM
TURBINE
S
M = 112.1
VACUUM
CONDENSER
1-1/2"Hg. 92°F

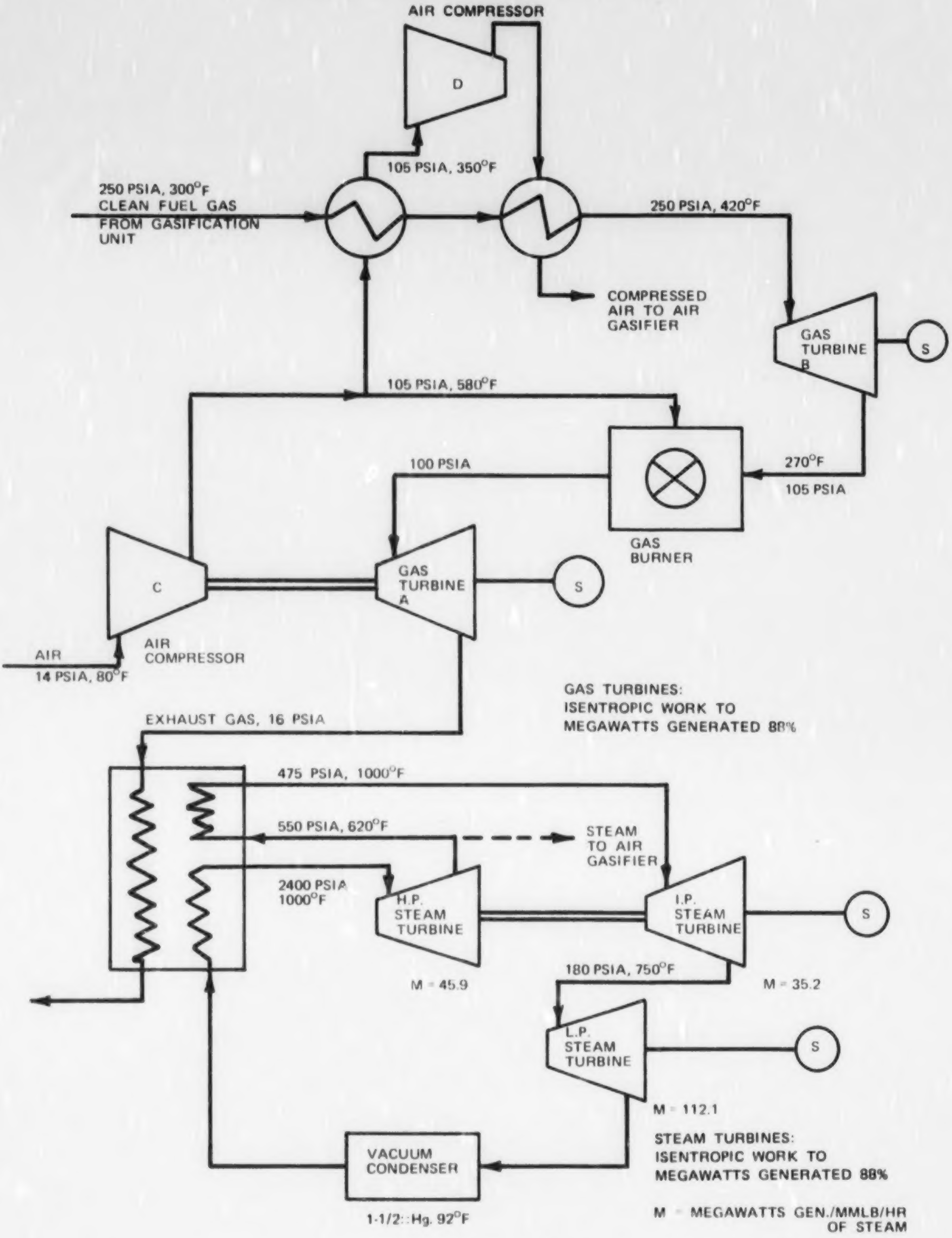
GAS TURBINES:
ISENTROPIC WORK TO
MEGAWATTS GENERATED 88%

M = MEGAWATTS/MMLB/HR OF STEAM

STEAM TURBINES:
ISENTROPIC WORK TO
MEGAWATTS GENERATED 88%

FIGURE 10.5

ADVANCED POWER CYCLE
POWER GENERATION SECTION (CYCLE 3)


AIR COMPRESSOR
D
105 PSIA, 350°F
250 PSIA, 300°F
CLEAN FUEL GAS
FROM GASIFICATION
UNIT
250 PSIA, 420°F
COMPRESSED
AIR TO AIR
GASIFIER
GAS
TURBINE
B
S
105 PSIA, 580°F
100 PSIA
270°F
105 PSIA
GAS
BURNER
C
GAS
TURBINE
A
S
AIR
14 PSIA, 80°F
AIR
COMPRESSOR
EXHAUST GAS, 16 PSIA
475 PSIA, 1000°F
550 PSIA, 620°F
STEAM
TO AIR
GASIFIER
2400 PSIA
1000°F
H.P.
STEAM
TURBINE
I.P.
STEAM
TURBINE
S
M = 45.9
M = 35.2
180 PSIA, 750°F
L.P.
STEAM
TURBINE
S
M = 112.1
VACUUM
CONDENSER
1-1/2" Hg. 92°F

GAS TURBINES:
ISENTROPIC WORK TO
MEGAWATTS GENERATED 88%

STEAM TURBINES:
ISENTROPIC WORK TO
MEGAWATTS GENERATED 88%

M = MEGAWATTS GEN./MMLB/HR
OF STEAM

FIGURE 10.6

ADVANCED POWER CYCLE
POWER GENERATION SECTION (CYCLE 4)

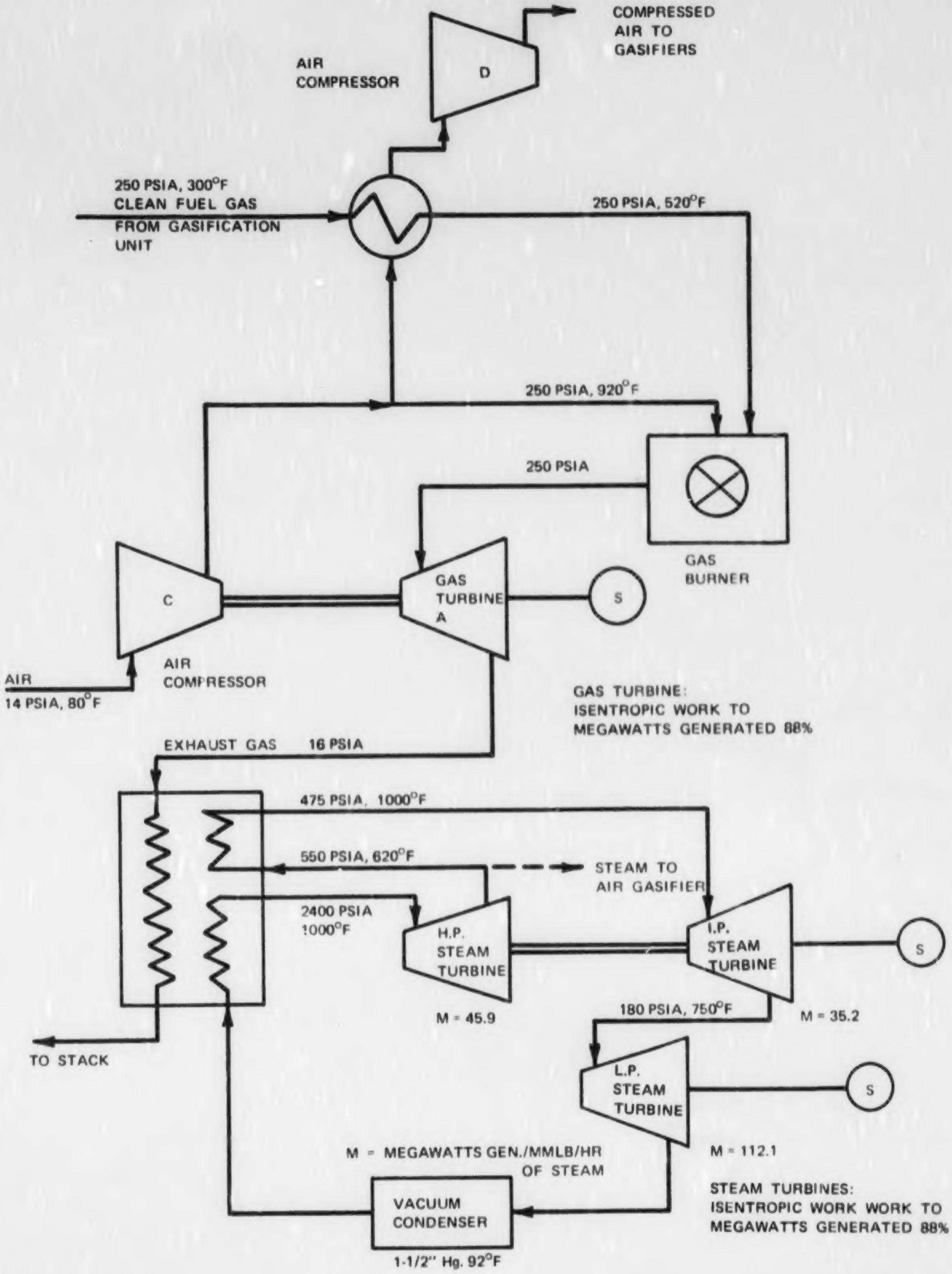
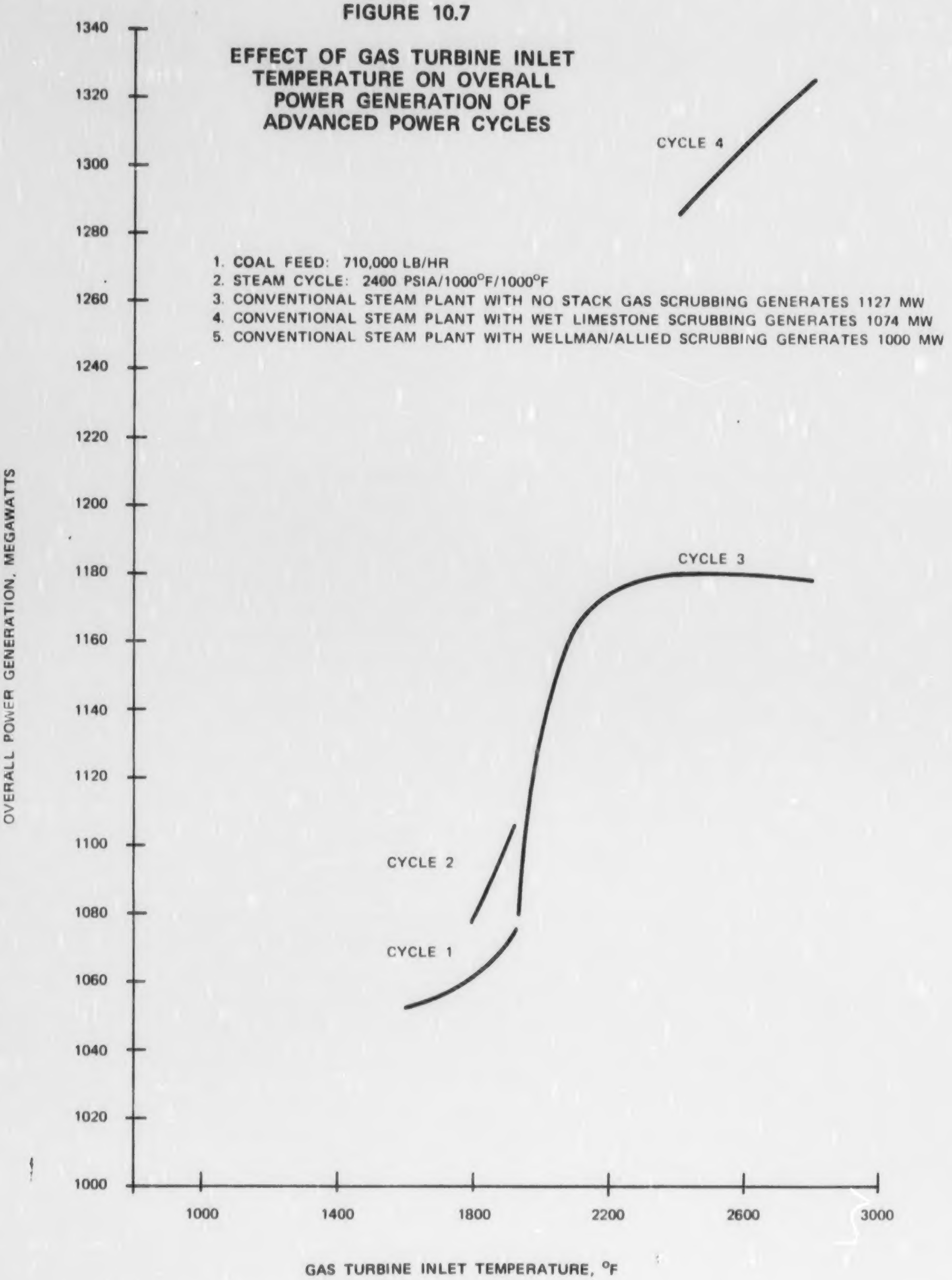

COMPRESSED AIR TO GASIFIERS
AIR COMPRESSOR
D
250 PSIA, 300°F CLEAN FUEL GAS FROM GASIFICATION UNIT
250 PSIA, 520°F
250 PSIA, 920°F
250 PSIA
GAS BURNER
C
GAS TURBINE A
S
AIR 14 PSIA, 80°F
AIR COMPRESSOR
GAS TURBINE: ISENTROPIC WORK TO MEGAWATTS GENERATED 88%
EXHAUST GAS
16 PSIA
475 PSIA, 1000°F
550 PSIA, 620°F
STEAM TO AIR GASIFIER
2400 PSIA 1000°F
H.P. STEAM TURBINE
I.P. STEAM TURBINE
S
M = 45.9
180 PSIA, 750°F
M = 35.2
TO STACK
L.P. STEAM TURBINE
S
M = MEGAWATTS GEN./MMLB/HR OF STEAM
M = 112.1
STEAM TURBINES: ISENTROPIC WORK WORK TO MEGAWATTS GENERATED 88%
VACUUM CONDENSER
1-1/2" Hg. 92°F

FIGURE 10.7

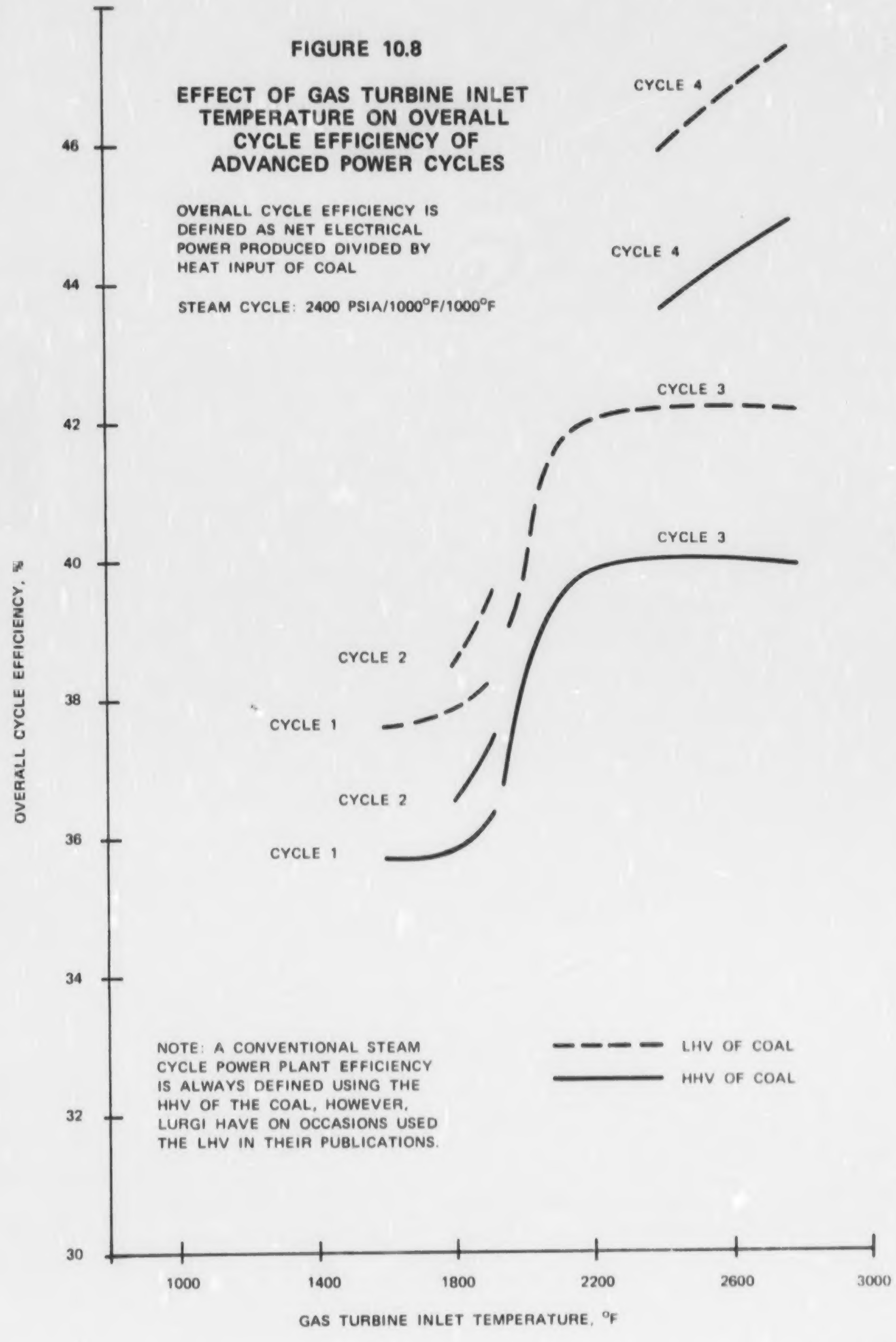
EFFECT OF GAS TURBINE INLET TEMPERATURE ON OVERALL POWER GENERATION OF ADVANCED POWER CYCLES



1. COAL FEED: 710,000 LB/HR
2. STEAM CYCLE: 2400 PSIA/1000°F/1000°F
3. CONVENTIONAL STEAM PLANT WITH NO STACK GAS SCRUBBING GENERATES 1127 MW
4. CONVENTIONAL STEAM PLANT WITH WET LIMESTONE SCRUBBING GENERATES 1074 MW
5. CONVENTIONAL STEAM PLANT WITH WELLMAN/ALLIED SCRUBBING GENERATES 1000 MW

FIGURE 10.8

EFFECT OF GAS TURBINE INLET TEMPERATURE ON OVERALL CYCLE EFFICIENCY OF ADVANCED POWER CYCLES



OVERALL CYCLE EFFICIENCY IS DEFINED AS NET ELECTRICAL POWER PRODUCED DIVIDED BY HEAT INPUT OF COAL

STEAM CYCLE: 2400 PSIA/1000°F/1000°F

NOTE: A CONVENTIONAL STEAM CYCLE POWER PLANT EFFICIENCY IS ALWAYS DEFINED USING THE HHV OF THE COAL, HOWEVER, LURGI HAVE ON OCCASIONS USED THE LHV IN THEIR PUBLICATIONS.

FIGURE 10.9

COMPRESSED AIR FLOW TO GAS BURNER VS. COMBUSTION TEMPERATURE

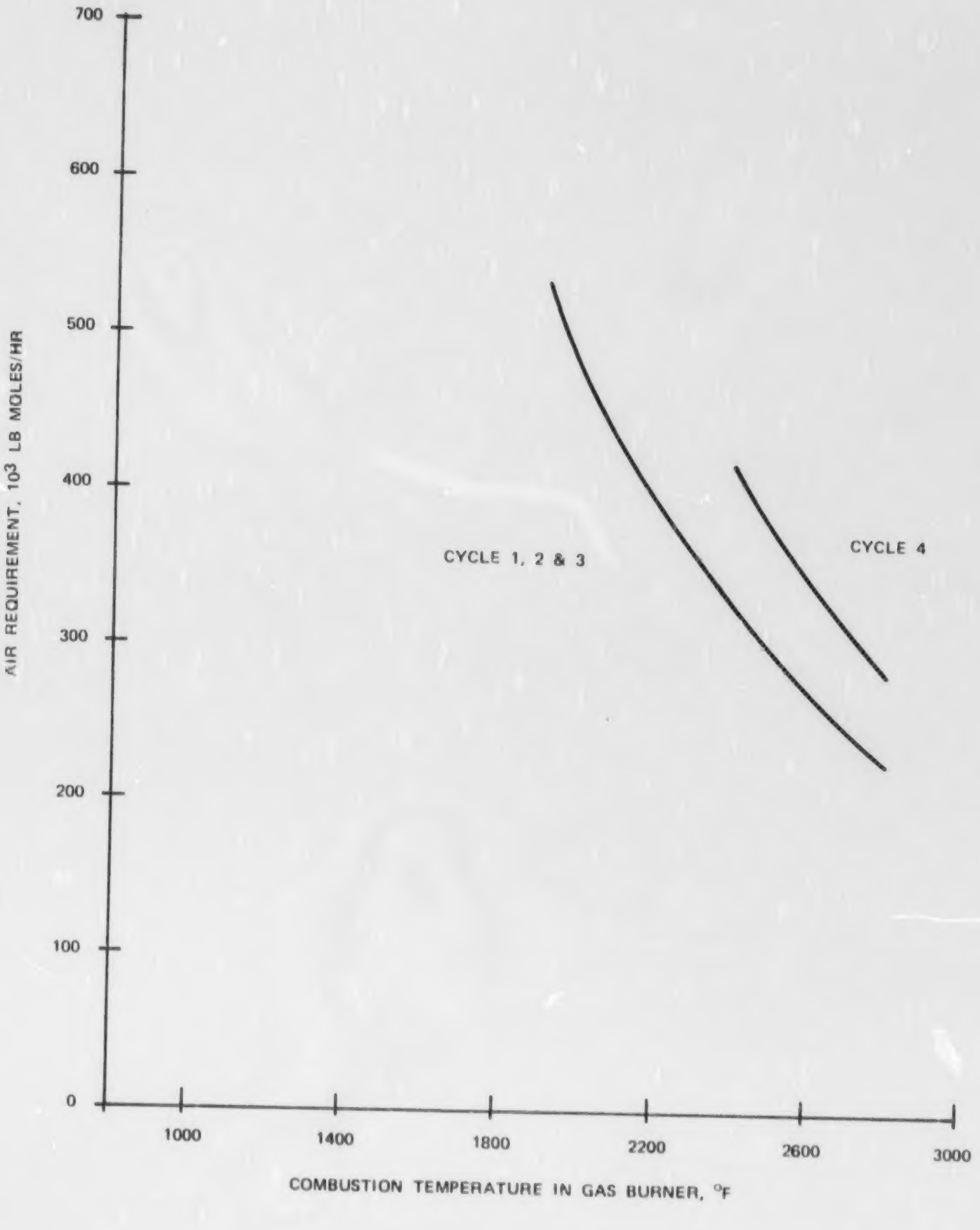


FIGURE 10.10

TURBINE EXHAUST GAS FLOW VS. GAS TURBINE INLET TEMPERATURE

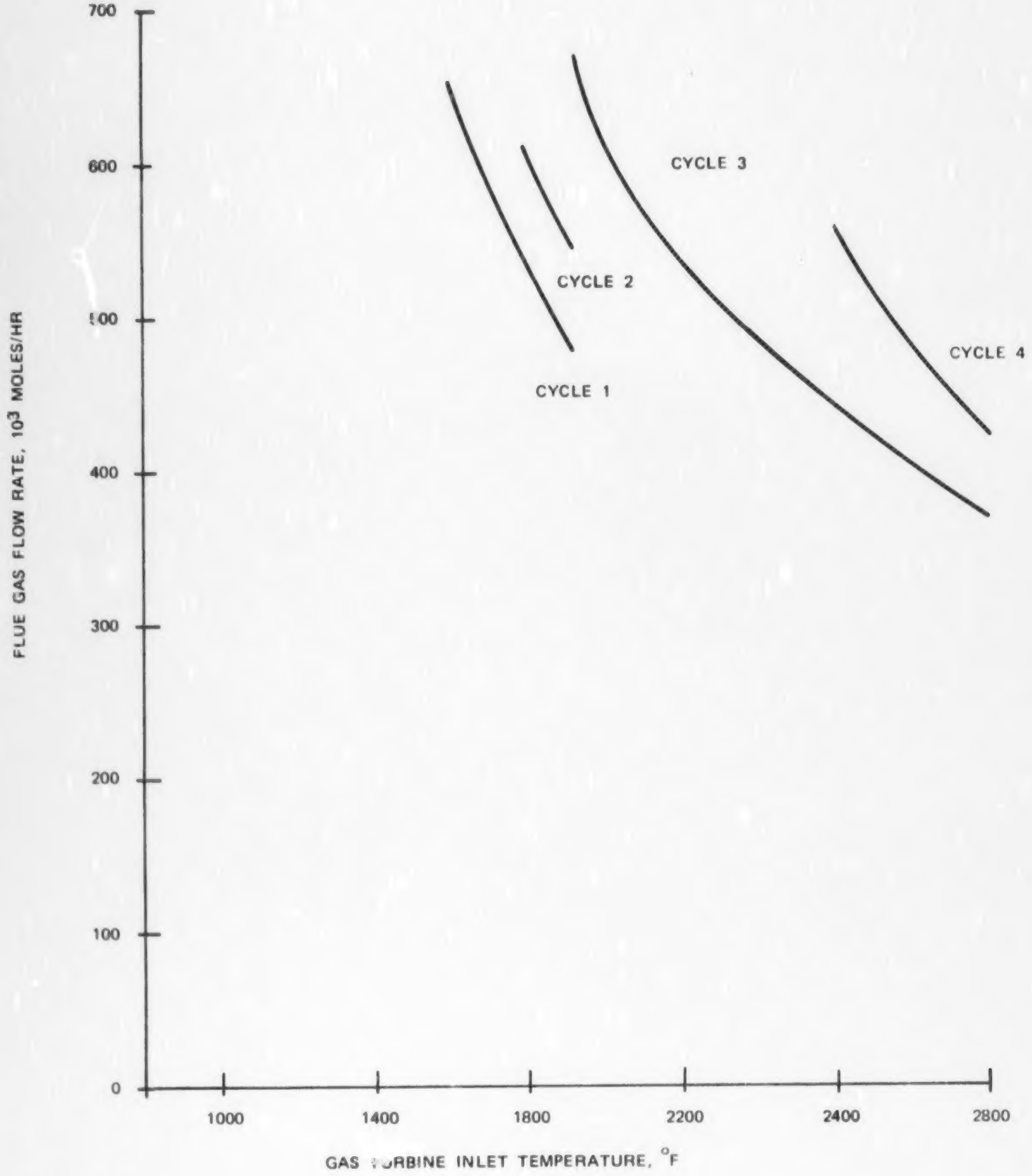


FIGURE 10.11

TURBINE POWER GENERATION AND AIR COMPRESSOR REQUIREMENTS FOR CYCLE 3

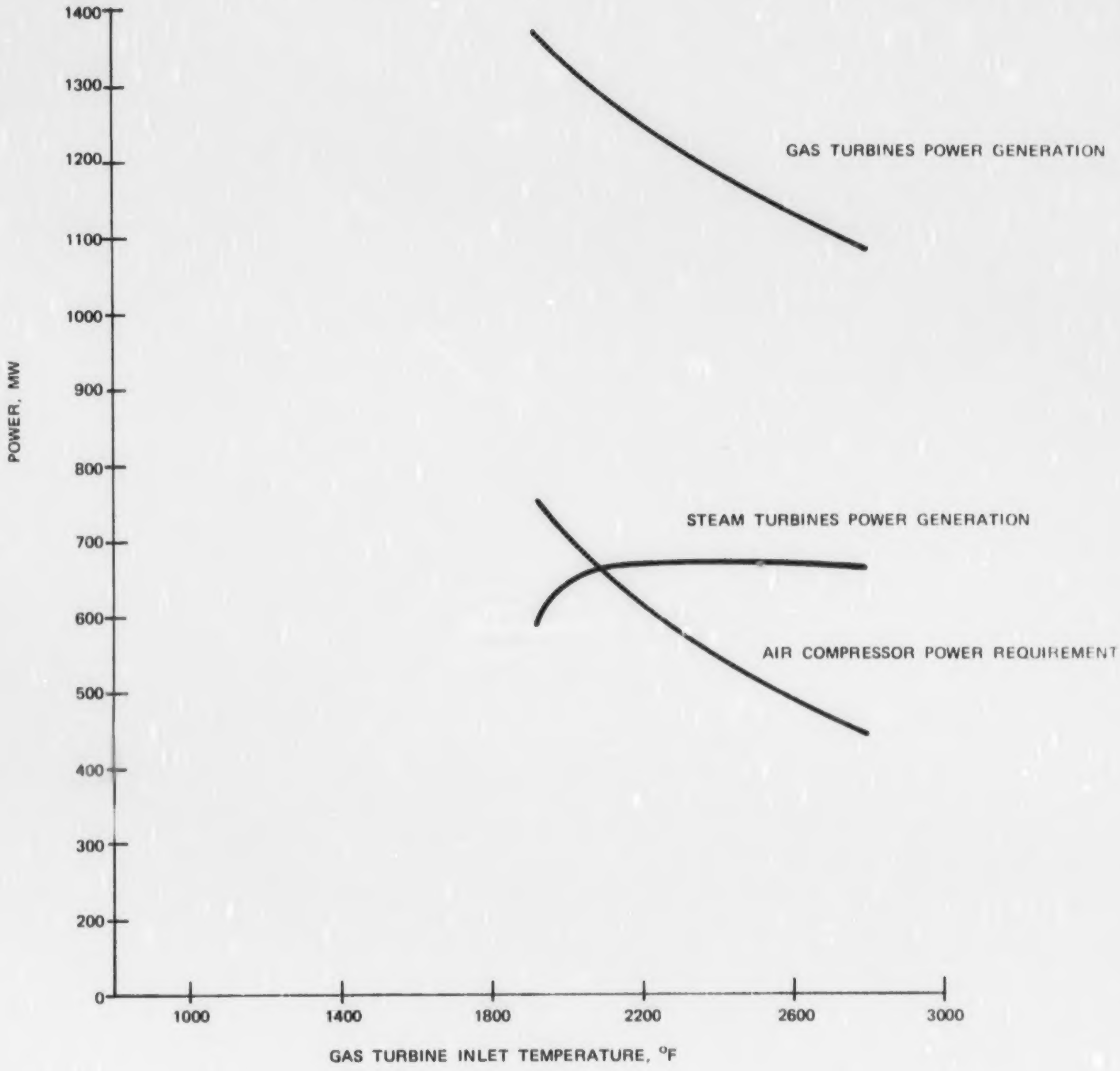
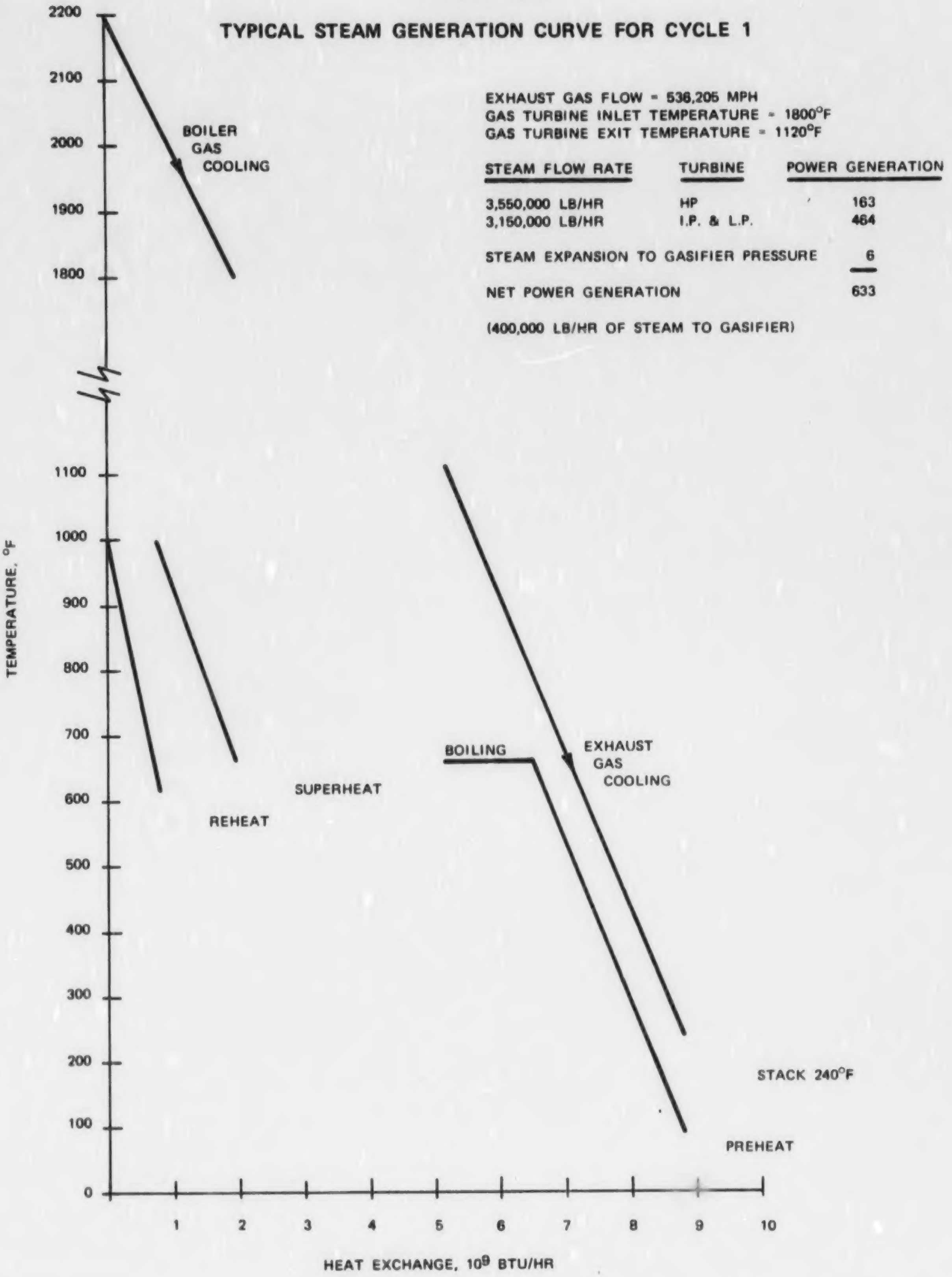


FIGURE 10.12

TYPICAL STEAM GENERATION CURVE FOR CYCLE 1



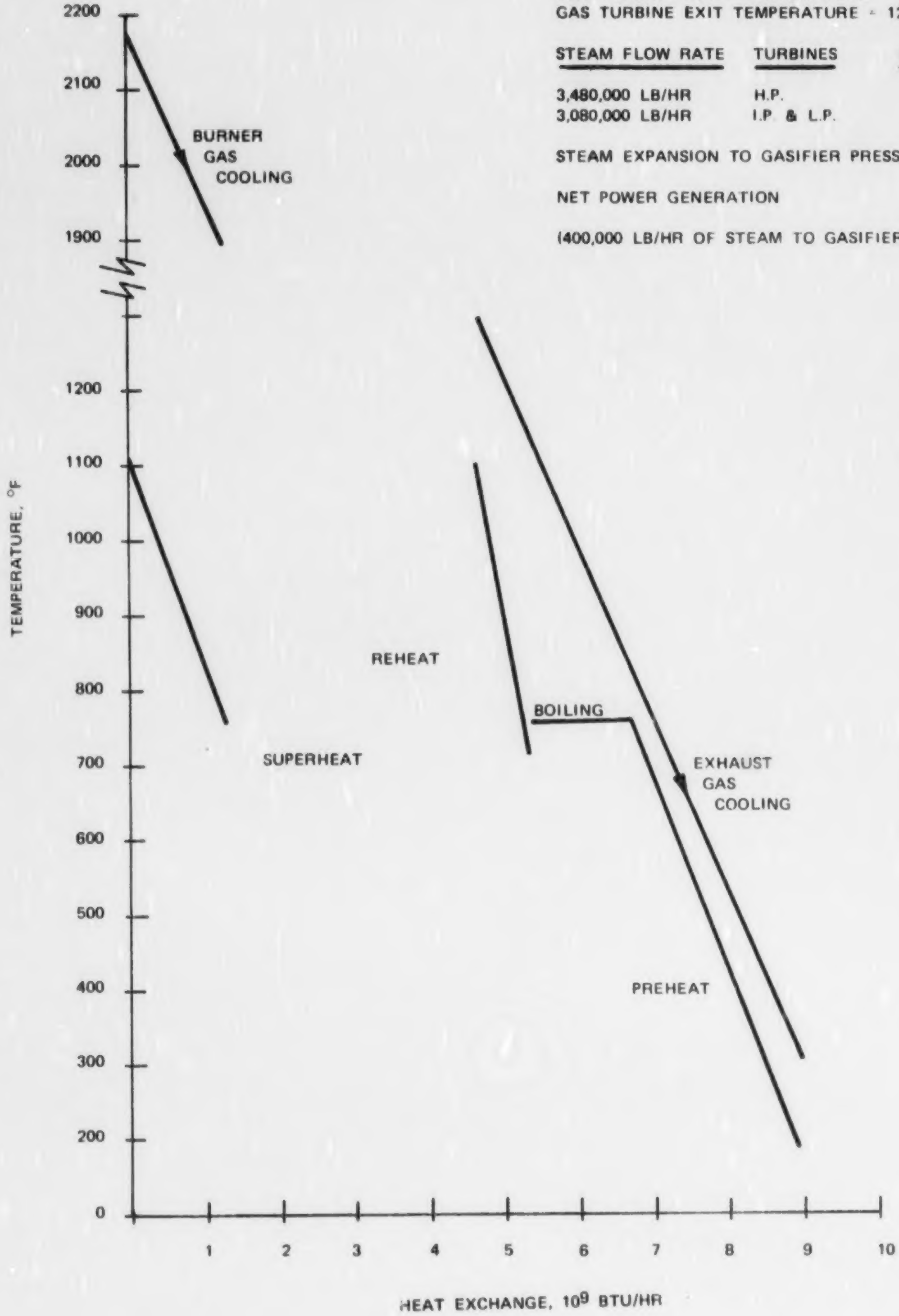
EXHAUST GAS FLOW = 536,205 MPH
GAS TURBINE INLET TEMPERATURE = 1800°F
GAS TURBINE EXIT TEMPERATURE = 1120°F

STEAM FLOW RATE	TURBINE	POWER GENERATION
3,550,000 LB/HR	HP	163
3,150,000 LB/HR	I.P. & L.P.	464
STEAM EXPANSION TO GASIFIER PRESSURE		6
NET POWER GENERATION		633

(400,000 LB/HR OF STEAM TO GASIFIER)

FIGURE 10.13

TYPICAL STEAM GENERATION CURVE FOR CYCLE 2



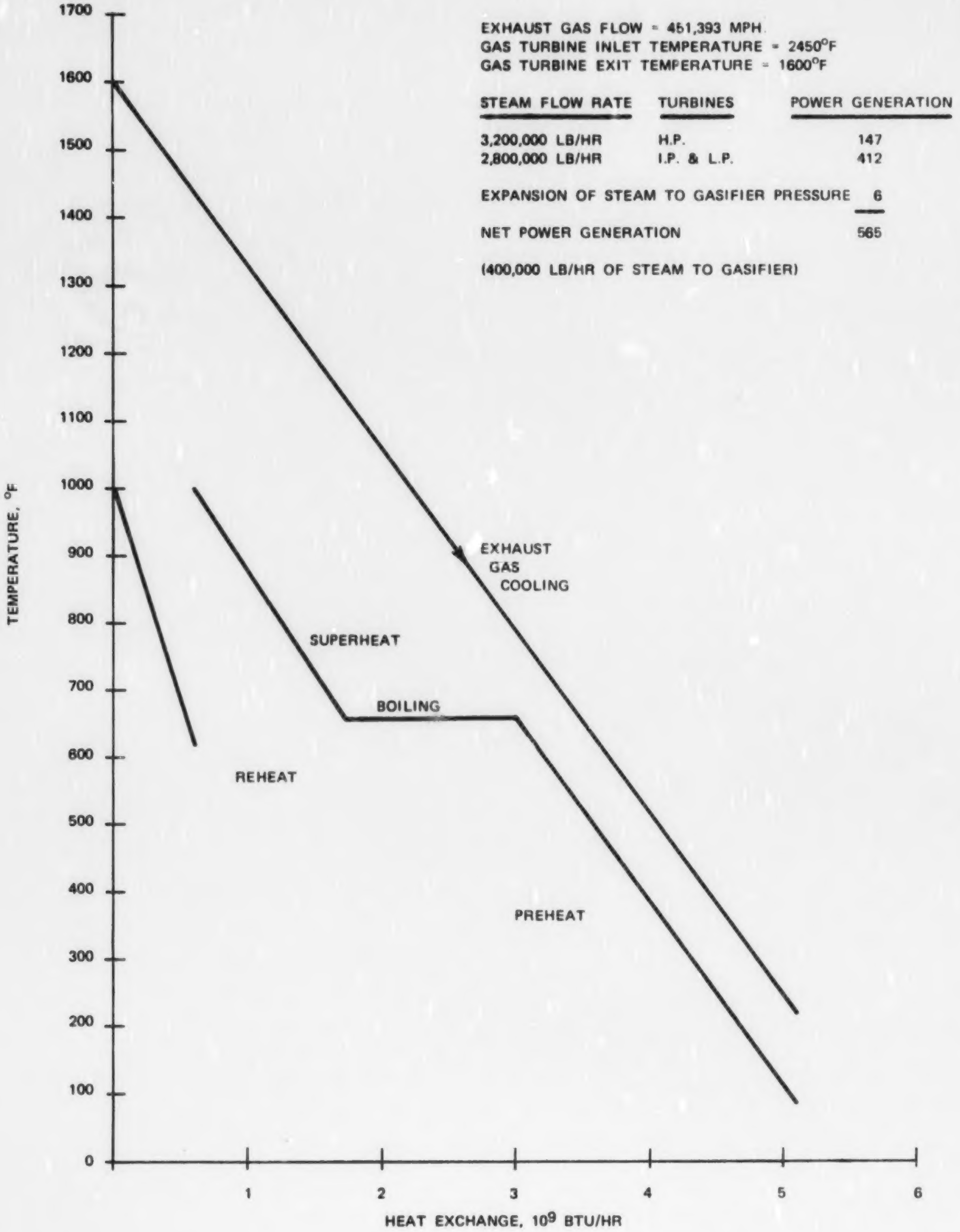
EXHAUST GAS FLOW = 546,571 MPH
GAS TURBINE INLET TEMPERATURE = 1920°F
GAS TURBINE EXIT TEMPERATURE = 1200°F

STEAM FLOW RATE	TURBINES	POWER GENERATION
3,480,000 LB/HR	H.P.	160
3,080,000 LB/HR	I.P. & L.P.	454
STEAM EXPANSION TO GASIFIER PRESSURE		6
NET POWER GENERATION		620

(400,000 LB/HR OF STEAM TO GASIFIER)

FIGURE 10.14

TYPICAL STEAM GENERATION CURVE FOR CYCLE 3 OR 4



EXHAUST GAS FLOW = 451,393 MPH
GAS TURBINE INLET TEMPERATURE = 2450°F
GAS TURBINE EXIT TEMPERATURE = 1600°F

STEAM FLOW RATE	TURBINES	POWER GENERATION
3,200,000 LB/HR	H.P.	147
2,800,000 LB/HR	I.P. & L.P.	412

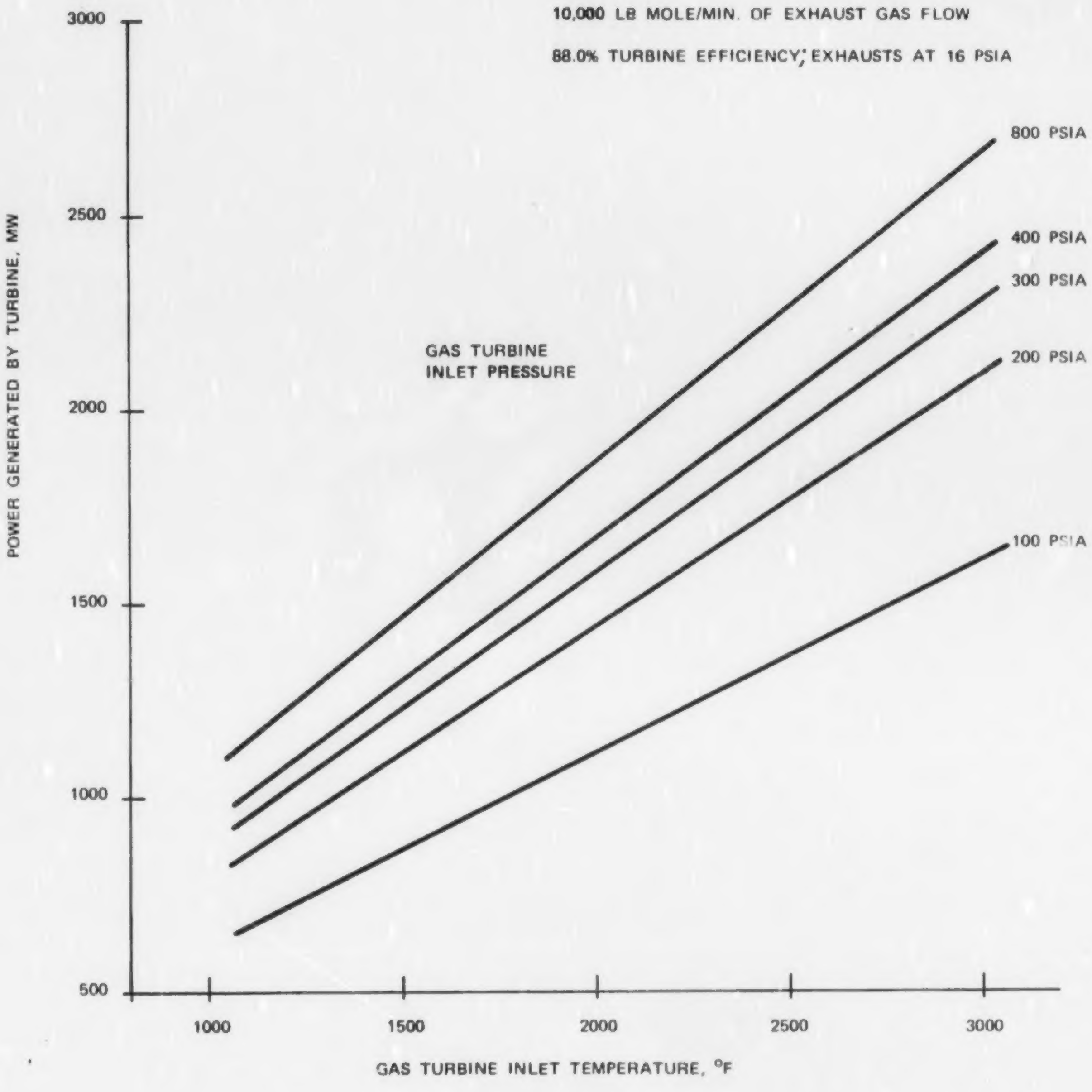
EXPANSION OF STEAM TO GASIFIER PRESSURE 6

NET POWER GENERATION 565

(400,000 LB/HR OF STEAM TO GASIFIER)

FIGURE 10.15

APPROXIMATE POWER GENERATION FOR GAS TURBINE A



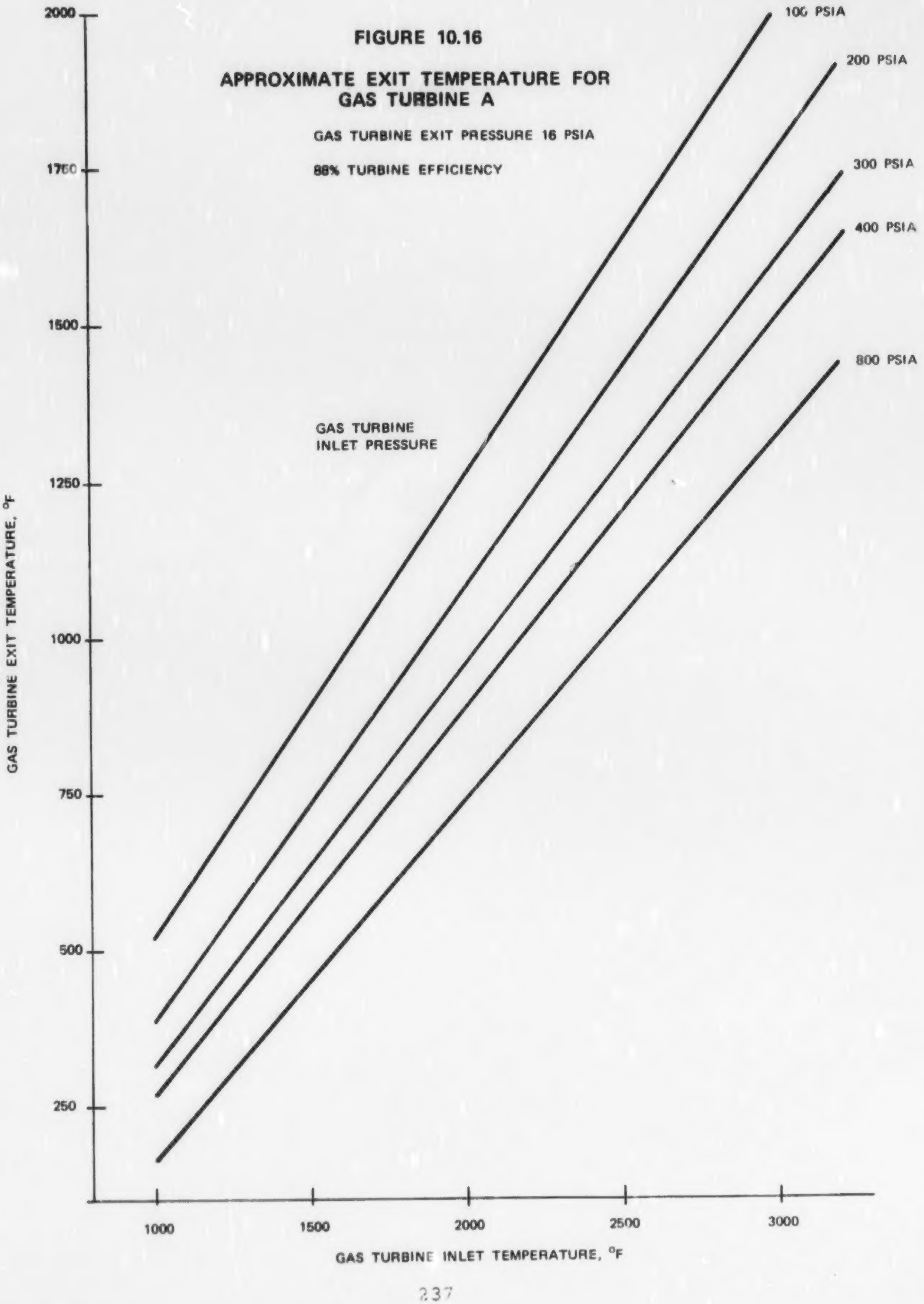


FIGURE 10.16

APPROXIMATE EXIT TEMPERATURE FOR GAS TURBINE A

GAS TURBINE EXIT PRESSURE 16 PSIA

88% TURBINE EFFICIENCY

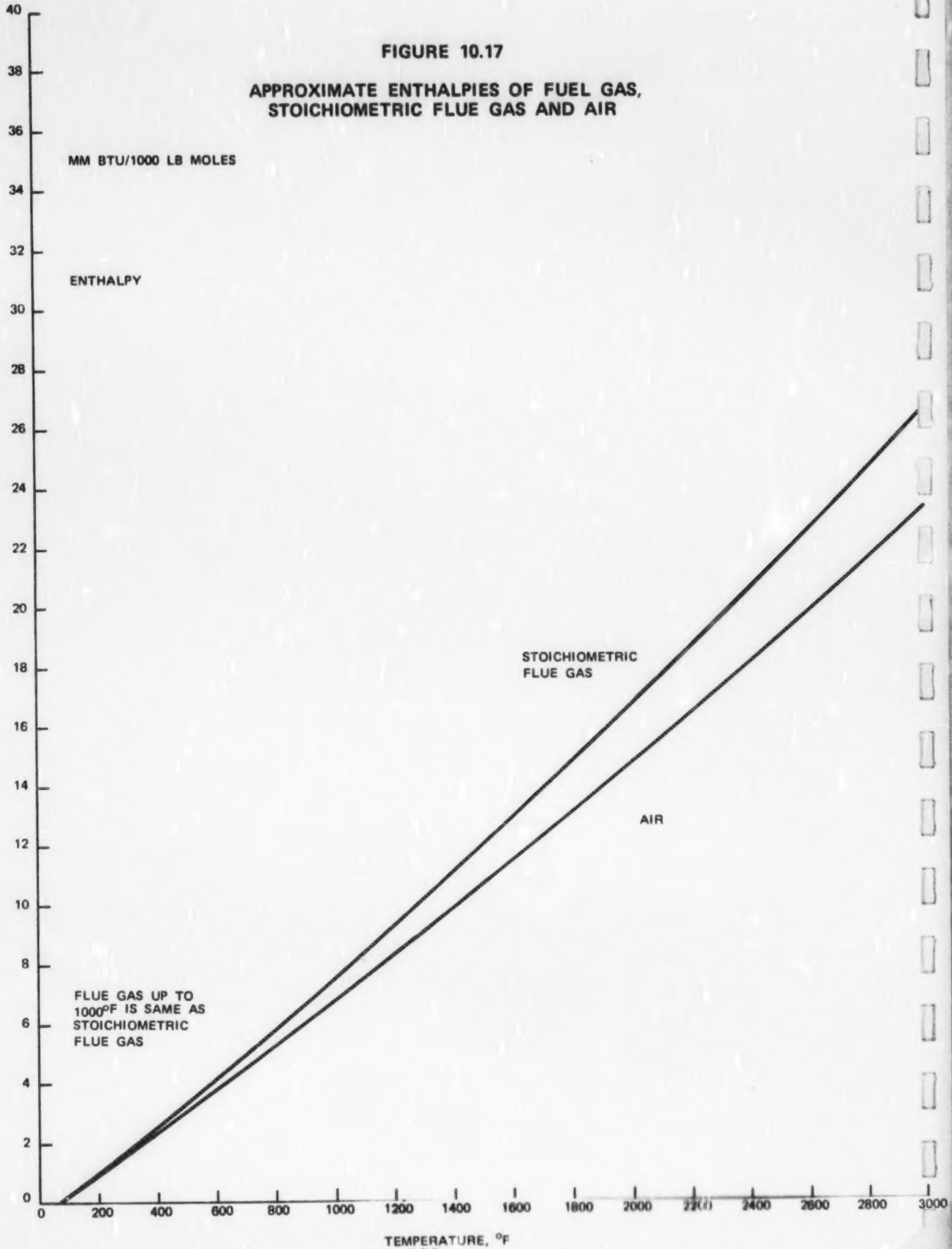
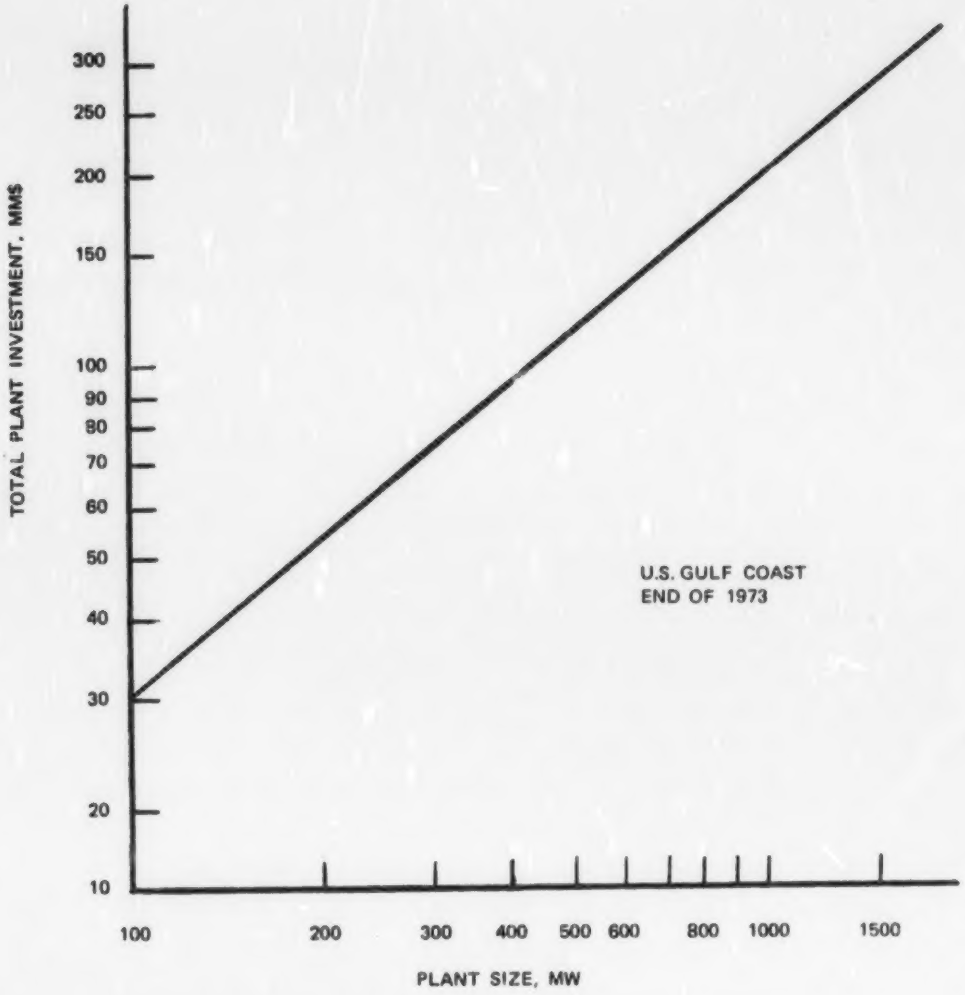


FIGURE 10.17

APPROXIMATE ENTHALPIES OF FUEL GAS, STOICHIOMETRIC FLUE GAS AND AIR

FIGURE 10.18

COST OF CONVENTIONAL COAL-FIRED STEAM POWER PLANT



11. PRESSURIZED FLUIDIZED BED STEAM GENERATOR WITH DRY DOLOMITE INJECTION FOR SO_2 REMOVAL

11.1. Process Appraisal

The fluidized bed combustion process, as evaluated and presented by Westinghouse in their report for EPA (41), was chosen as the basis for the development of this model and their design was examined by MWK Research & Engineering Development Department. The process includes a regenerative dolomite system for SO_2 removal. Material and heat balances prepared by Westinghouse were in agreement with ours, and we have accepted Westinghouse's statement of reactions taking place under the conditions specified by them. However, in some areas their design was found to be unrealistic.

The regenerative system involves several large units and compressors. The design of the solids handling systems would present enormous difficulties. In addition, experimental results show that the regenerative efficiency falls off markedly with the number of cycles. The costs for a dolomite regeneration system appear to be unrealistic, and we do not feel convinced by Westinghouse's low figures for an item which has yet to be constructed commercially. However, in view of the decision to drop the regenerative scheme and concentrate on once through dolomite design, the costs and problems of a regenerative system are academic.

To ensure adequate turn down facility for the pressurized fluidized bed steam generator, it is necessary to use four boiler modules, which means that the concept becomes less desirable for small plant sizes and industrial applications.

11.2. Process Description

The vertical pressure vessel of the fluidized boiler has four exchangers, mounted one above another. The preheater section is at the bottom. Above this are the evaporator, superheater, and reheater beds. The pulverized coal is introduced at the bottom of the bed, and about six times the stoichiometric amount of dolomite (limestone) enters at the top of the bed to react with sulfur in the coal. About 10% excess air is supplied to the fluidized bed, giving a superficial velocity 6-9 ft/sec and a bed temperature of 1750°F. The entrained solids are recycled to a carbon burn-up cell in the combustor vessel itself, which operates at 79% excess air at 2000°F. The overall combustor unit takes 15% excess air and about 1% of the carbon is eventually lost without combustion.

The flue gases from the CBC are passed through the second stage of a particulate separator, before entering the gas turbines. Flue gases enter the gas turbine at 1600°F and leave at 900°F, generating 490 megawatt. The sensible heat of the turbine effluent is used to preheat the feed water in two stages. The flue gases enter the stack at 200°F.

SO_2 evolved during combustion reacts with lime to form $CaSO_4$:

$$CaCO_3 \rightarrow CaO + CO_2$$

$$CaO + SO_2 + 1/2\ O_2 \rightarrow CaSO_4$$

The regeneration flowsheet is as described by Westinghouse.

Sulfated dolomite from the fluidized bed boiler is converted back to carbonate by reducing to calcium sulfide and subsequent regeneration with steam and CO_2.

$$CaSO_4 + 4CO \rightarrow CaS + 4CO_2$$

$$CaSO_4 + 4H_2 \rightarrow CaS + 4H_2O$$

$$CaS + H_2O + CO_2 \rightarrow CaCO_3 + H_2S$$

$CaSO_4$ is reduced to CaS by producer gas in a reactor at about 1500°F and 135 psia. The producer gas is generated by oxidizing coal with an air-steam mixture. Steam controls the gas temperature and provides H_2 for reduction.

Feed gas for the second stage regenerator is obtained by purifying CO_2 from boiler flue gas. A slip-steam from stack gas is compressed to 135 psia and absorbed in regenerable hot carbonate. CO_2 is then stripped from the carbonate and cooled to 200°F before recompression to 180 psia. It is then fed to the H_2S generator. Rich H_2S leaving the generator at 1100°F and 165 psia is expanded through a turbine to 2 atmosphere before sending to the Claus unit.

11.3 Conclusions

A fluidized bed gives better heat transfer and more uniform temperature distribution. Surface area requirements are reduced by 60-70% due to the high heat transfer rate. Pressurized fluidized bed operation has certain other advantages to conventional boilers:

1. It can burn low grade, high sulfur coals efficiently while conforming to stringent air pollution control regulations.
2. NO_x emissions are reduced substantially.
3. Cycle efficiencies of 35-39% can be achieved with dry dolomite injection for SO_2 removal.

Although this design gives a higher gas-steam combined cycle efficiency than a conventional steam cycle with stack gas scrubbing, there are certain limitations and problem areas for pressurized industrial boilers which need development work.

The major areas which need further consideration are:

1. More stringent particulate removal is needed before the gas turbine.
2. High temperature and high pressure piping, valving and ducting, a particulate removal system, plus coal and dolomite feeding systems have to be used. This is inherently expensive.
3. Turn down is the big problem in pressurized fluidized bed operation. As shown by Westinghouse data, even by using four modules there is a discontinuity in the turn down.
4. Nothing definite has been established about regenerative efficiency of dolomite with time.

11.4 Addendum

After our review was completed, Westinghouse Corporation issued a second set of reports on the evaluation of the fluidized bed combustion process (EPA 650/2-73-D48 a, b, c, and d, December, 1973). These reports contain information on sorbent requirements for a once-through sulfur removal system, regeneration system costs, regeneration system potential, turn down capabilities and development requirements. The readers are encouraged to refer to this set of reports for the latest information.

TABLE 11.1

GAS STREAMS

STREAM	TEMP °F	PRESS PSIA	FLOW MPH/HR	MOLE %						
				H_2	H_2O	CO	CO_2	N_2	O_2	H_2S
G1	80	14.7	287,500	-	-	-	-	79	21	-
G2	700	150	278,000	-	-	-	-	79	21	-
G3	700	150	9,500	-	-	-	-	79	21	-
G4	640	150	750	-	100	-	-	-	-	-
G5	1500	135	13,173	8.0	10.7	16.3	7.4	57.1	-	0.5
G6	1500	116	10,630	1.4	21.7	0.5	5.5	70.9	-	-
G7	1600	150	295,250	-	8.8	-	14.3	74.3	2.6	-
G8	300	135	34,120	-	8.8	-	14.3	74.3	2.6	-
G9	200	19	10,050	-	63.7	-	36.3	-	-	-
G10	1100	165	7,642	-	73.3	-	16.2	-	-	10.5
G11	1000	2400/475	295,278	-	100	-	-	-	-	-

TABLE 11.2

SOLID STREAMS

STREAM	DESCRIPTION	FLOWRATE lb/hr	COMPOSITION WT%
S1	Total coal feed to plant	710,000 DAF	C 78, H 5.5, O 11.0, S 4.0, N1.5 DAFB
S2	Coal feed to combustor	662,000 DAF	C 78, H 5.5, O 11.0, S 4.0, N1.5 DAFB
S3	Coal feed to gas generator	48,000 DAF	C 78, H 5.5, O 11.0, S 4.0, N1.5 DAFB
S4	Dolomite make up	96,000	80 $CaCO_3$, 20 $MgCO_3$
S5	Regenerated stone to combustor	700,000	60 $CaCO_3$, 20 CaO, 20 MgO
S6	Sulfated dolomite to reducer	630,000	16 $CaSO_4$, 64 CaO, 20 MgO
S7	Spent stone purge	81,000	60 $CaCO_3$, 20 CaO, 20 MgO

TABLE 11.3

Power Generation of an FBC Combined Cycle Plant

GENERATION	MEGAWATTS
Net Steam Cycle	916
Gas Turbine	517
Reducer Reactor Effluent Turbine	16
H_2S Generator Reactor Effluent Turbine	10
	1459

REQUIREMENTS	
Air Compressor (including air to producer gas generator)	366
CO_2 Compressor	12
Flue Gas Slipstream Compressor	36
Auxiliary Power Other Than Steam Cycle	10
Equivalent Power of Steam to Producer Gas Generator and CO_2 Stripper Reboiler	15
	439

Net Power Generation = 1020 megawatts.

Overall Cycle Efficiency = 34.4% (HHV of coal)

NOTES: 1). Plant uses regenerative dolomite system for sulfur control.

2). Total plant feed is 710,000 lb/hr DAF bituminous coal containing 4% sulfur

TABLE 11.4

Heat to Steam Cycle

Heat Losses From Combustor	% of Coal HHV
Radiation and Convection	1.8
Incomplete Combustion	1.5
Heats of Calcining Reactions etc.	0.4
Heat to bring Dolomite up to 1700°F	0.3
Heat Loss by Hot Ash	0.4
Heat Loss by Transferences between Combustor and Regenerator	0.6
Total	5.0

HEAT IN WITH RAW MATERIALS	MMBTU/HR
HHV of coal to combustor	9,400
Enthalpy of air @ 700°F	1,250
	10,650
Less heat losses above	-470
Net useful heat	10,180
HEAT OUT WITH COMBUSTED GASES	
Sensible @ 1600°F	3,697
Latent Ht. of water	488
	4,185

HEAT INTO STEAM CYCLE = 10,180 - 4185

= 5,995 MMBTU/HR

TABLE 11.5

Cost of a 1000 Net Megawatt FBC Combined Cycle Plant

POWER GENERATION	TOTAL PLANT INVESTMENT, TPI (M$)
Coal handling and injection system	15,000
Pressurized combustor/boiler	50,000
Steam-turbo generator and condensers	44,000
Air compressor/gas turbine/generator	25,000
Land and Structures	14,000
Accessory electrical equipment	8,000
Transmission plant	4,000
	160,000
REGENERATION OF DOLOMITE	
Producer gas generator	2,000
$CaSO_4$ reducer unit	5,000
H_2S generator unit	5,000
Sulfur Recovery	5,000
CO_2 absorber/stripper unit	4,000
Compressor and turbines	4,000
Other offsites and solids transportation	5,000
	30,000

Total for dolomite regeneration and power station is $190 million.

NOTES: 1). Plant uses a regenerative dolomite system for sulfur control.

2). Costs are Gulf Coast, end of 1973.

FIGURE 11.1

FLOW DIAGRAM FOR FLUIDIZED PRESSURIZED COAL COMBUSTOR POWER GENERATOR

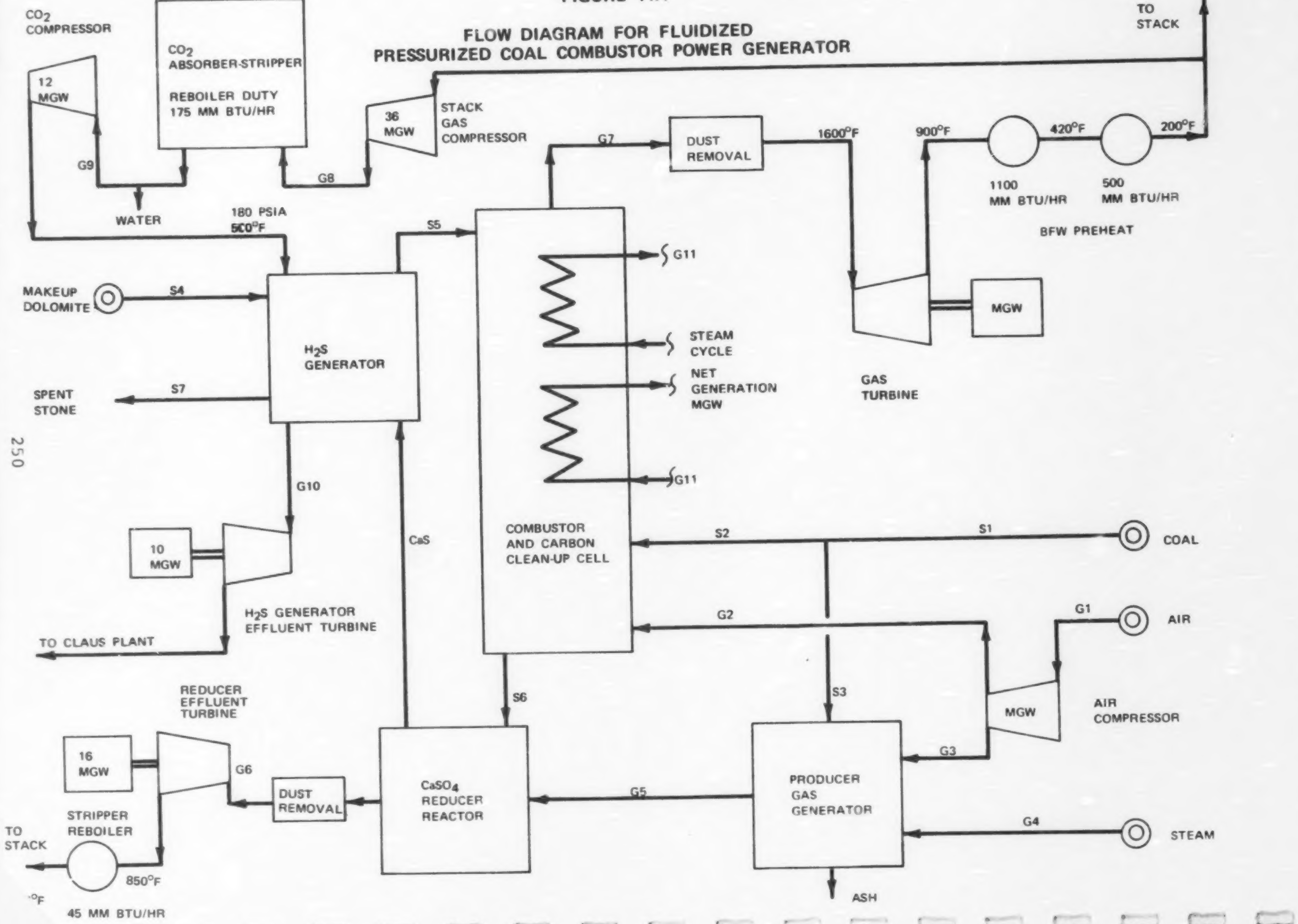
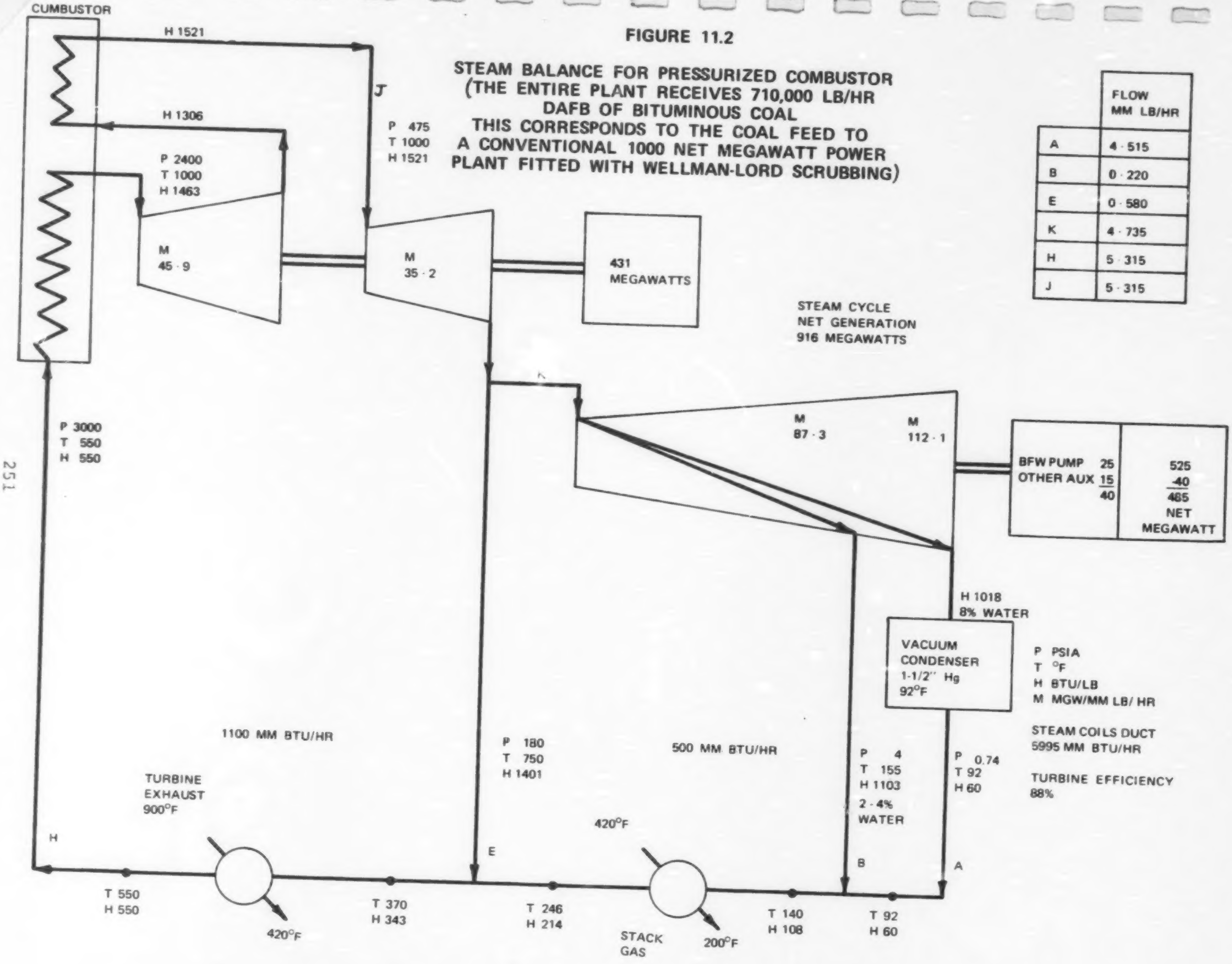

CO_2 COMPRESSOR
12 MGW
G9
WATER
CO_2 ABSORBER-STRIPPER
REBOILER DUTY 175 MM BTU/HR
G8
36 MGW
STACK GAS COMPRESSOR
TO STACK
180 PSIA
500°F
S5
G7
DUST REMOVAL
1600°F
900°F
420°F
200°F
1100 MM BTU/HR
500 MM BTU/HR
BFW PREHEAT
MAKEUP DOLOMITE
S4
H_2S GENERATOR
G11
STEAM CYCLE
NET GENERATION MGW
G11
MGW
GAS TURBINE
SPENT STONE
S7
G10
10 MGW
H_2S GENERATOR EFFLUENT TURBINE
TO CLAUS PLANT
CaS
COMBUSTOR AND CARBON CLEAN-UP CELL
S2
S1
COAL
G2
G1
AIR
S6
S3
MGW
AIR COMPRESSOR
REDUCER EFFLUENT TURBINE
16 MGW
G6
DUST REMOVAL
$CaSO_4$ REDUCER REACTOR
G5
G3
PRODUCER GAS GENERATOR
G4
STEAM
STRIPPER REBOILER
TO STACK
850°F
°F
45 MM BTU/HR
ASH

FIGURE 11.2

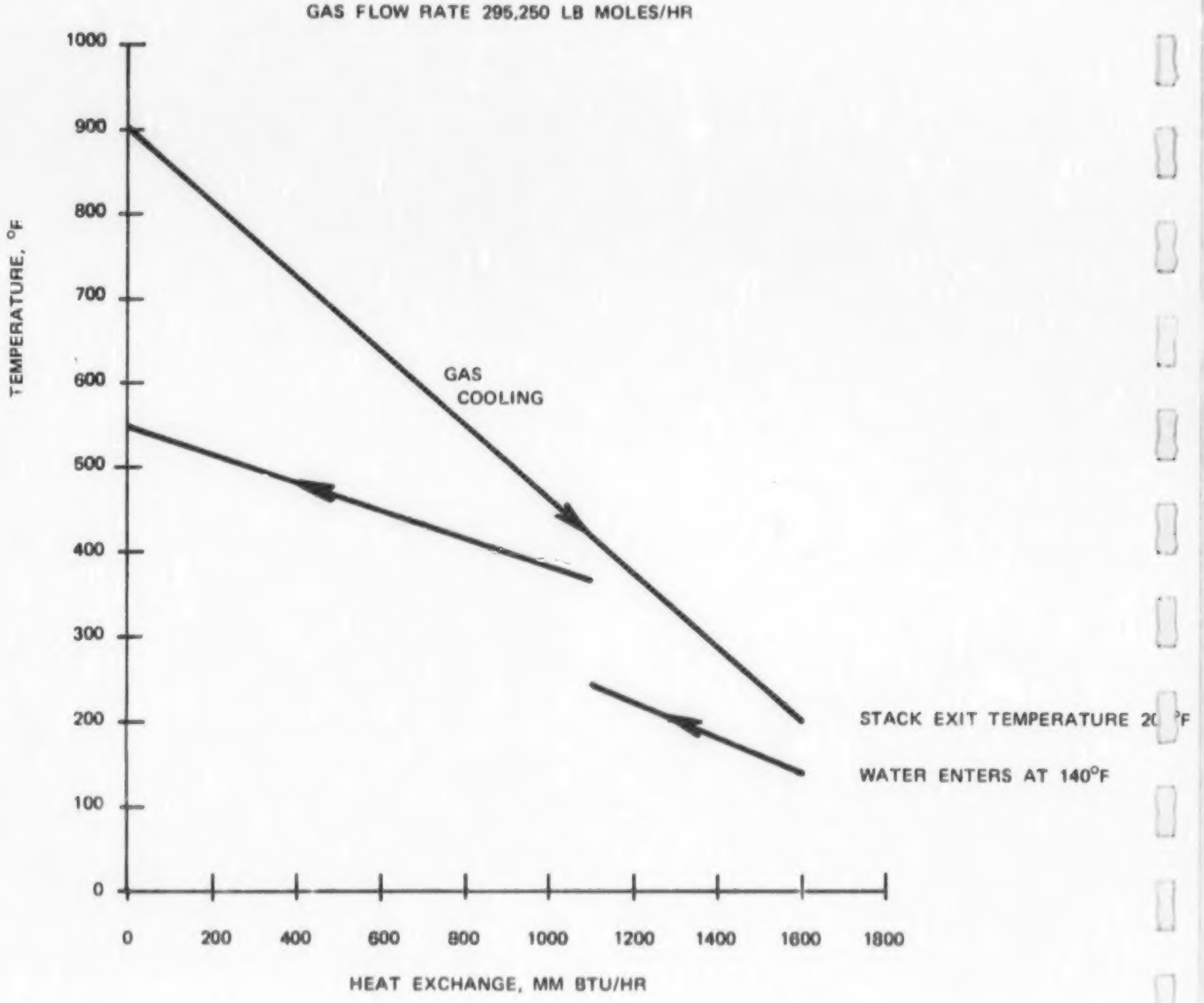
STEAM BALANCE FOR PRESSURIZED COMBUSTOR
(THE ENTIRE PLANT RECEIVES 710,000 LB/HR
DAFB OF BITUMINOUS COAL
THIS CORRESPONDS TO THE COAL FEED TO
A CONVENTIONAL 1000 NET MEGAWATT POWER
PLANT FITTED WITH WELLMAN-LORD SCRUBBING)


CUMBUSTOR
H 1521
J
H 1306
P 2400
T 1000
H 1463
P 475
T 1000
H 1521
M
45 · 9
M
35 · 2
431
MEGAWATTS
STEAM CYCLE
NET GENERATION
916 MEGAWATTS
K
M
87 · 3
M
112 · 1
BFW PUMP 25
OTHER AUX 15
40
525
-40
485
NET
MEGAWATT
H 1018
8% WATER
VACUUM
CONDENSER
1-1/2" Hg
92°F
P PSIA
T °F
H BTU/LB
M MGW/MM LB/ HR
STEAM COILS DUCT
5995 MM BTU/HR
TURBINE EFFICIENCY
88%
P 3000
T 550
H 550
1100 MM BTU/HR
TURBINE
EXHAUST
900°F
P 180
T 750
H 1401
500 MM BTU/HR
420°F
P 4
T 155
H 1103
2 · 4%
WATER
P 0.74
T 92
H 60
H
E
B
A
T 550
H 550
420°F
T 370
H 343
T 246
H 214
STACK
GAS
200°F
T 140
H 108
T 92
H 60

	FLOW MM LB/HR
A	4 · 515
B	0 · 220
E	0 · 580
K	4 · 735
H	5 · 315
J	5 · 315

FIGURE 11.3

GAS TURBINE EXHAUST COOLING CURVE



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42. "Evaluation of the Fluidized Bed Combustion Process", Prepared for Office of Research & Development, U. S. Environmental Protection Agency by Westinghouse Laboratories, Pittsburg, Penn. (Contract No. 68-02-0217) Volume I, II, & III, December 1973.

13. APPENDICES

APPENDIX A

General Cost Model Derivations

Interest During Construction (IDC)

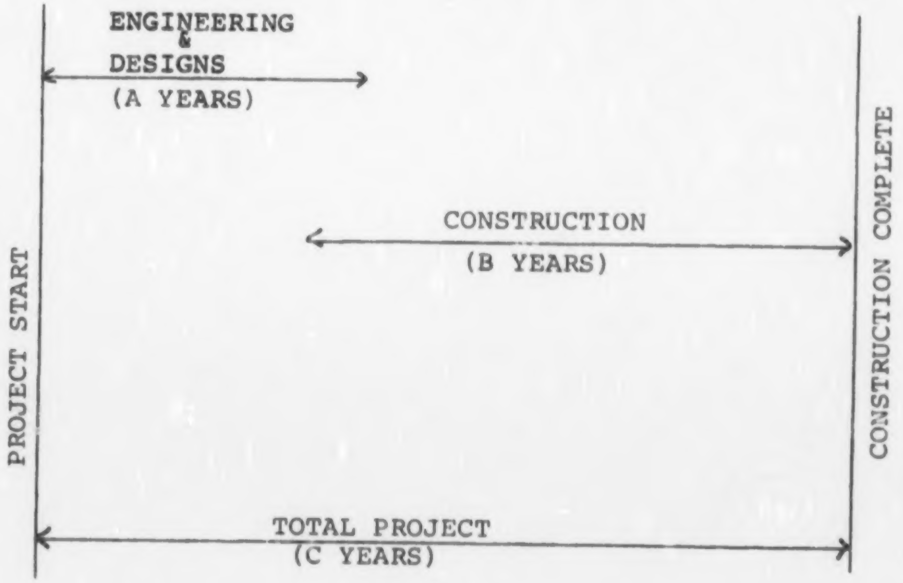
Interest during construction represents the cost of interim financing of a project during the design and construction period. If the project is financed on borrowed capital, it is a real cost to the company. If corporate funds are used, it is an internal charge equivalent to the income which would have been obtained if the capital had been used for short-term investment at normal commercial interest rates.

The total interest during construction can be obtained from the following general equation:

interest = capital x interest rate x time

where "capital" is the total construction cost of the plant, i.e., the total plant investment (TPI). The time period referred to in the equation is a function of the project schedule as it affects cash flow.

A typical project schedule can be represented as follows:


ENGINEERING & DESIGNS
(A YEARS)
PROJECT START
CONSTRUCTION
(B YEARS)
CONSTRUCTION COMPLETE
TOTAL PROJECT
(C YEARS)

APPENDIX A (cont'd)

From the General Cost Model, the cost of engineering and design is normally about 10% of the total plant investment while construction costs about 90%. Assuming a uniform cash flow during the engineering and design phase (A yrs), the interest charge would be approximately equal to the interest rate applied to the entire engineering and design cost over one-half the time period (For interest purposes, a uniform cash flow is roughly equivalent to a single cash flow at the mid-point of the time period). When the engineering and design phase is completed, interest continues to accumulate until the end of the construction period (C-A years). Thus, the total interest charge on the engineering and design costs, I_1, is:

$$\begin{aligned} I_1 &= 0.1 \text{ TPI} \times i \times \frac{A}{2} + 0.1 \text{ TPI} \times i \times (C-A) \\ &= 0.1 \text{ TPI} \times i \left[\frac{A}{2} + (C-A)\right] \\ &= 0.1 \text{ TPI} \times i \left(C-\frac{A}{2}\right) \end{aligned}$$

The interest on construction costs, I_2, would be approximately equal to the interest rate applied to the entire cost over one-half the construction period.

Thus:

$$I_2 = 0.9 \text{ TPI} \times i \times \frac{B}{2}$$

The total interest during construction, IDC, is therefore given by:

$$\begin{aligned} \text{IDC} &= I_1 + I_2 \\ &= 0.1 \text{ TPI} \times i \left(C-\frac{A}{2}\right) + 0.9 \text{ TPI} \times i \times \frac{B}{2} \\ &= \text{TPI} \times i \left[0.1 \left(C-\frac{A}{2}\right) + 0.9 \left(\frac{B}{2}\right)\right] \end{aligned}$$

In order to obtain values for IDC for use in the General Cost Model, typical project schedules have been assumed as follows:

	A (Yrs)	B (Yrs)	C (Yrs)
Stack gas scrubbing	1.5	2.5	3.0
SNG, SRC, power plants	2.5	3.5	5.0

Assuming an interest rate of 9%/year and substituting in the preceding equation, we obtain:

IDC = 0.12 TPI	Stack gas scrubbing
IDC = 0.18 TPI	SNG, SRC, power plants

Interest on Debt and Return on Equity (AIC)

The calculation of AIC is based on the utility method used by the Synthetic Gas-Coal Task Force of the FPC National Gas Survey and illustrated in their final report (13). The method assumes that the total capital required, TCR, is split into a percentage debt (borrowed capital) and a percentage equity (owned capital). The debt portion is charged at the commercial interest rate while the equity is charged at some desired net rate of return. Depreciation covers return of capital for both the debt and equity portions of TCR. Interest on debt and return on equity are calculated over the life of the plant and the average yearly value is expressed as a fraction of TCR.

At any given time, the book or asset value of a plant equals its original cost (TCR) less the total accrued depreciation. This quantity, which represents the debt still outstanding plus the equity capital yet to be recovered, is the rate base upon which interest on debt and return on equity are calculated. For any given year, the average rate base, $(BR)_i$, equals TCR less the accrued depreciation at the mid-point of the year. Thus, $(AIC)_i$, the annual interest on debt plus return on equity for year i, is calculated as follows:

$$(AIC)_i = (BR)_i \times \text{fraction debt} \times \text{fraction annual interest rate}$$
$$+ (BR)_i \times \text{fraction equity} \times \text{fraction annual net rate of return}$$

The average value of AIC over the life of the plant, expressed as a fraction of TCR, is:

$$AIC = \frac{\sum_{1}^{N} (AIC)_i}{N} \Big/ TCR$$

This is the equation which is used in the General Cost Model.

Typical values were assumed for fraction debt, fraction equity, interest rate, and net rate of return to obtain a numerical value for AIC for use in the General Cost Model. A sample calculation of AIC, based on these assumed values, is shown in Table A-1 on a year-by-year basis.

Note that since straight-line depreciation has been used, the rate base, and therefore the return on rate base, decreases linearly with time. Thus, AIC alternately could be calculated by using the average rate base over the plant life. This average base is equal to TCR less one-half of the total depreciation:

$$(BR)_{AVG} = TCR - \frac{1}{2}(TCR-WKC)$$
$$= \frac{1}{2}(TCR + WKC)$$

Substituting the assumed values of TCR and WKC from Table A-1:

$$(BR)_{AVG} = \frac{1}{2}(100{,}000 + 3{,}000)$$
$$= \$51{,}500M$$

Using the assumed values in Table A-1 for percent debt, percent equity, interest rate, and rate of return, AIC can be calculated as follows:

$$\begin{aligned} \text{AIC} &= (0.75 \times 0.09 + 0.25 \times 0.15)\ (\text{BR})_{\text{AVG}} \\ &= 0.105 \times 51{,}500 \\ &= \$5407\ \text{M/Year} \end{aligned}$$

It should be noted that plant life does not enter into the calculation. Thus, the expression for AIC used in the General Cost Model is valid for all types of plants (stack gas scrubbing units, SNG plants, power plants, etc.).

Federal Income Tax (AFT)

From Table A-1, the average net (after tax) return on equity is $1,931M. Expressed as a fraction of TCR, this is:

$$\begin{aligned} \text{net return} &= \frac{1{,}931}{100{,}000} \cdot \text{TCR} \\ &= 0.0193\ \text{TCR} \end{aligned}$$

Assuming an income tax rate of 48%, the net return represents 52% of the gross (pre-tax) return. The federal income tax is therefore given by:

$$\begin{aligned} \text{AFT} &= \frac{\text{net income}}{0.52} \times 0.48 \\ &= 0.0193\ \text{TCR} \times \frac{0.48}{0.52} \\ &= 0.018\ \text{TCR} \end{aligned}$$

TABLE A-1

SAMPLE CALCULATION OF AIC

Assumptions

Total Capital Required (TCR)	= $100,000 M
Working Capital (WKC)	= $ 3,000 M
Plant Life	= 20 years
Debt	= 75% of TCR
Equity	= 25% of TCR
Interest rate on debt	= 9%/year
Required net rate of return on equity	= 15%/year

(1) YEAR	(2) AVG. RATE BASE, BR (TCR-accrued deprec. @ mid-point of year)	(3) INTEREST ON DEBT (0.75x0.09xBR)	(4) NET RETURN ON EQUITY (0.25x0.15xBR)	(5) RETURN ON RATE BASE (3 + 4)	(6) DEPRECIATION (0.05 x [TCR-WKC])
0	$100,000 M	-	-	-	-
1	97,575 M	$ 6,586 M	$ 3,659 M	$ 10,245 M	$ 4,850 M
2	92,725	6,259	3,477	9,736	4,850
3	87,875	5,932	3,295	9,227	4,850
4	83,025	5,604	3,113	8,717	4,850
5	78,175	5,277	2,932	8,209	4,850
6	73,325	4,949	2,750	7,699	4,850
7	68,475	4,622	2,568	7,190	4,850
8	63,625	4,295	2,386	6,681	4,850
9	58,775	3,967	2,204	6,171	4,850
10	53,925	3,640	2,022	5,662	4,850
11	49,075	3,313	1,840	5,153	4,850
12	44,225	2,985	1,658	4,643	4,850
13	39,375	2,658	1,477	4,135	4,850
14	34,525	2,330	1,295	3,625	4,850
15	29,675	2,003	1,113	3,116	4,850
16	24,825	1,676	931	2,607	4,850
17	19,975	1,348	749	2,097	4,850
18	15,125	1,021	567	1,588	4,850
19	10,275	694	385	1,079	4,850
20	5,425	366	203	569	4,850
20 year total		69,525	38,624	108,149	97,000
20 year average		3,476	1,931	5,407	

$$AIC = \frac{5,407}{100,000} \times TCR = 0.054\ TCR$$

APPENDIX B

WET LIMESTONE PROCESS-CATALYTIC INC. ESTIMATE, 500 MEGAWATT EQUIPMENT COSTS (MATERIAL & SUBCONTRACTS)

(Inflation index from mid 72 to the end of 73 = 1.08)

Scrubbing System

	Mid 1972 $	End of 1973 $
4 Venturi Scrubbers (380,000 ACFM at the inlet to each)	253,000	273,000
4 Two-stage TCA's	840,000	908,000
4 Sumps for Venturis & TCA's	254,000	274,000
4 Horizontal Chevron Entrainment Separators	531,000	574,000
4 Venturi Tanks	85,000	92,000
4 TCA Tanks	121,000	131,000
4 Venturi Tank Agitators	26,400	28,500
4 TCA Tank Agitators	31,000	33,500
4 Sets of Venturi Recirculation Pumps	47,600	51,500
4 Sets of TCA Recirculation Pumps	91,000	98,300
Ammonia Injection System	10,000	10,800
Entrainment Separator Recir. Tank	28,500	30,800
Entrainment Separator Recir. Pumps	30,900	33,400
	2,349,400	2,538,800

Flue Gas Reheat and Discharge

	Mid 1972 $	End of 1973 $
Ductwork including dampers etc.	1,003,200	1,085,000
4 I.D. Fans and Motors (360,000 ACFM, 37 inch w.g., 3000 BHP)	332,000	359,000
4 Reheater Burner Units	156,000	169,000
Fuel Oil Tankage and Loading Pump	69,270	75,000
	1,560,470	1,688,000

APPENDIX B (con't)

WET LIMESTONE PROCESS-CATALYTIC INC. ESTIMATE, 500 MEGAWATT EQUIPMENT COSTS (MATERIAL & SUBCONTRACTS)

(Inflation index from mid 72 to the end of 73 = 1.08)

Limestone Handling and Slurry Preparation

	Mid 1972 $	End of 1973 $
Limestone Silo Conveyor & Stockpile Feeder	57,050	61,700
Limestone Storage Silo with 3 Cones	76,000	82,000
3 Limestone Weigh Feeders	21,400	23,100
3 Tube Mill Wet Grinders	550,000	595,000
Tube Mill Air Compressor	12,500	13,500
Tube Mill Surge Tank	1,500	1,600
Limestone Slurry Transfer Pumps	2,200	2,400
Limestone Slurry Hold Tank	46,300	50,000
Limestone Slurry Tank Agititator	26,900	29,100
Limestone Feed Pumps	2,900	3,200
	796,750	861,600

Waste Disposal

	Mid 1972 $	End of 1973 $
Sumps and Tankage	6,150	6,700
Pumps & Drives (inc. process water pumps)	31,840	34,500
Separating Pond (250 acres 50 ft. deep) Pond Located in Cincinnati.)	3,694,000	4,000,000
	3,731,990	4,041,200

APPENDIX C

WET LIMESTONE PROCESS-CATALYTIC INC. ESTIMATE, LABOR AND MATERIAL FACTORS

MAJOR EQUIPMENT COSTS

	CATALYTIC INC Part I P65 $		This Report $
Material	2,925,300	EC chemical process	3,947,900
Subcontract	1,819,300	ES solid handling	796,700
Total	4,744,600	Total	4,744,600

FIELD LABOR COSTS (Cincinnati with location factor = 1.53)

	$		$
		LC chemical process	2,363,000
		LS solid handling	212,000
Total	2,575,000	Total	2,575,000

OTHER MATERIAL COSTS

(Piping, instrumentation, electrical, civil etc.)

	$		$
Total Material	6,218,700	MC chemical process	3,225,700
Maj. Equip. Mat.	2,925,300	MS solid handling	67,700
Other Material	3,293,400	Total	3,293,400

LC = 0.60 EC

LS = 0.27 ES

MC = 0.82 EC

MS = 0.09 ES

APPENDIX D

A SUMMARY OF THE COMMENTS MADE DURING AND AFTER THE PRESENTATION BY MWK TO EPA, DECEMBER 13, 1973

1. There was agreement with the $1.2 to $1.4/MMBtu cost given for SNG, but surprise by some to find that this cost did not represent almost any location at a higher coal cost. There was disbelief that a contract to sell coal at anything like $3/ton would ever be signed. It was stressed by MWK that the utility industry in 1970 paid an average of over $7/ton.

2. Mr. K. Janes expressed the view that costs for solvent refined coal production while higher than those given in the Stearns-Roger estimate are still on the low side.

MWK agreed with this and pointed out that the accuracy of the plant investment was only within ± 30% and most probably the figure would be as underestimate if an actual commercial design were ever costed.

3. There was considerable interest in the Lurgi combined cycle presentation and a general agreement with the cycle efficiencies given in the presentation. The feeling expressed by a few people was that Lurgi was not as far advanced in the field of gasification and combined power cycles as Lurgi publications say. It was stated by Mr. P. Spaite that there have been considerable technical difficulties with Lurgi's Steag 165 megawatt unit and the gasification unit was not as reliable as was needed for a combined cycle. It was felt that reliability and better control of coal feed rather than cost reduction were the main reasons for development of other gasifier designs. The opinion that the hot carbonate purification unit was not proved in this service was expressed by at least two people. It was pointed out by MWK that Benfield Corporation felt confident of their design

and that Lurgi also had a hot carbonate design. Several people wondered why the costs for the hot carbonate unit were much lower than those for the SNG purification unit and the Wellman/Allied stack gas scrubbing units. It was pointed out by MWK that the purpose of the Rectisol unit used in the SNG plant was much different from that of the hot carbonate unit. The rectisol unit had to remove almost all of the CO_2 in the gas stream compared to the hot carbonate's 30%. The main difference was that the acceptable sulfur level for the methanator catalyst was 0.1 ppm whereas the hot carbonate left as much as 500 ppm H_2S. The Rectisol unit was more complicated and required a refrigeration unit. The coal feed to the combined cycle Lurgi unit was approximately 70% of that to the Lurgi SNG plant. Again, the function of the Wellman/Allied unit was much different in that it handled a flue gas with a flowrate approximately 2.5 times more (1000 megawatt plant) than that of the fuel gas from the low Btu Lurgi unit. The costs of the hot carbonate unit should also be added to the cost of the sulfur recovery unit before even a broad comparison could be made to the Wellman/Allied plant.

4. There was interest in the relative recovered energies of the SNG unit, the SRC unit and the power plants. MWK stated that the HHV of the product SNG was 58% of the HHV of the input coal and the corresponding figure was 79% for the SRC units compared to the best power plant efficiency of 40%. There was strong feeling by several people present that direct comparison was misleading and the relative forms of the energy had to be taken into account. SNG required power for transportation and electrical heating devices were more efficient than gas heating devices. There was agreement to some extent by MWK, but it was stated that there was a need to study transportation costs of the various alternatives and establish the final efficiency after consumption. MWK also pointed out that the transportation of low sulfur Wyoming coal to the Eastern states ought to be a more sensible use of the coal rather than gasification.

APPENDIX D (cont'd)

5. There was agreement with the conclusions expressed about the Westinghouse fluidized, pressurized combustor with regenerative dolomite sulfur removal. It was felt by MWK that regeneration was not technically possible to any consistent level by a commercial plant of the Westinghouse design. On top of this the costs for a finallized design could be many million dollars more than those presented. This view was substantiated by a statement by EPA that Westinghouse had now dropped the idea of regeneration and interest was now directed towards the pressurized combustor with sulfur removal either by a once through throwaway dolomite system or by stack gas scrubbing. It was stated by MWK that the stack gas scrubbing alternative would produce overall cycle efficiencies about the same as a conventional power plant with stack gas scrubbing. The only possible advantage would be one of reduced cost of the pressurized fluidized boiler over the conventional boiler. It is obviously in Westinghouse's best interest to establish whether this is in fact so at the earliest possible time.

It is likely that the once through dolomite alternative would have a higher overall cycle efficiency than the conventional plant with Wellman/Allied stack gas scrubbing; it may possibly be better than the conventional plant with Wet Limestone Scrubbing. A critical factor may be how many times the stiochiometric flow of dolomite is required. Waste disposal may be even more expensive than with the wet limestone process. The main areas for establishing concrete facts at an early date are the cost of the equivalent sized pressurized boiler compared to the conventional unit and the efficiency of sulfur removal at a stated dolomite flowrate.

END